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Costs for Advanced Coal Combustion Technologies

Radian Corp., Research Triangle Park, NC

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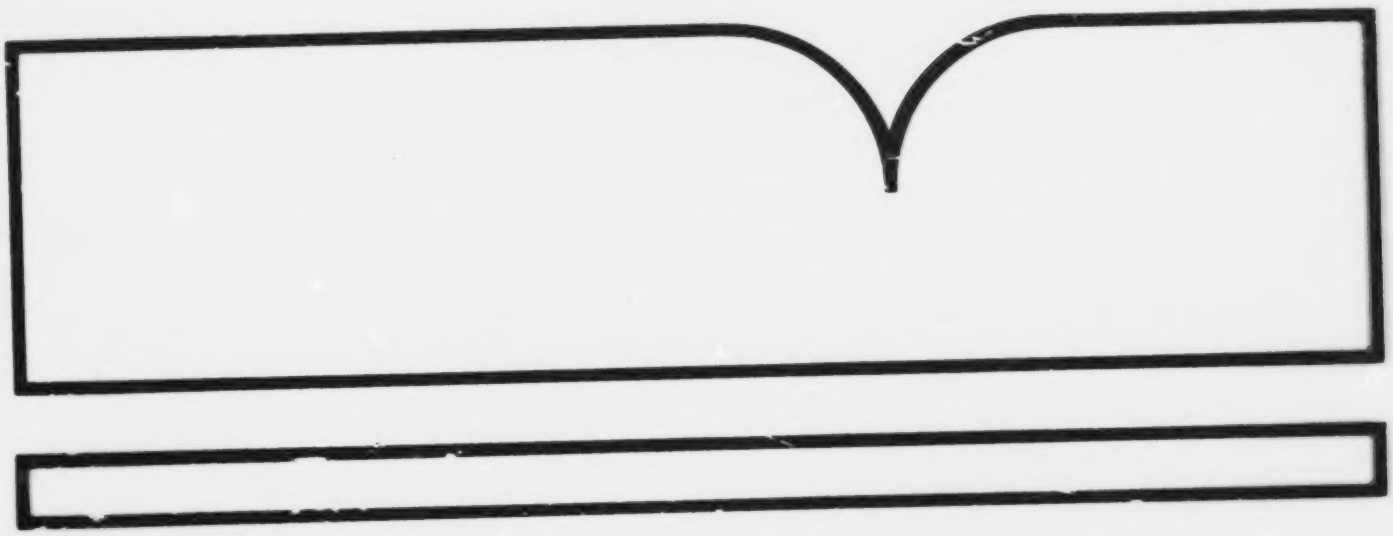

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COSTS FOR ADVANCED COAL COMBUSTION TECHNOLOGIES

by

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16. ABSTRACT The report gives results of an evaluation of the development status of advanced coal combustion technologies and discusses the preparation of performance and economic models for their application to electric utility plants. The technologies addressed were atmospheric fluidized bed combustion (AFBC), pressurized fluidized bed combustion (PFBC), and integrated gasification combined cycle (IGCC). The development status was also reviewed for pulverized-coal-fired (PC) boilers incorporating supercritical steam cycles. Although advanced combustion technologies are attractive due to decreased SO2 and NOx emission characteristics and potentially higher generating efficiencies, full commercial readiness does not appear feasible before the mid to late 1990s. Capital cost estimates for a new plant at 500 MW ranged from $1,250 to $1,910/kW for advanced technologies and from $1,380 to $1,810/kW for conventional systems. Capital cost estimates for PFBC (turbocharged cycle) and conventional plants with add-on SO2 and NOx controls were within 4% of the median cost ($1,580/kW) of all technologies evaluated. AFBC costs averaged 12% less than this median, while IGCC costs averaged 11% above the median. Potential capital cost savings for repowering an existing plant versus constructing a new facility at the same final capacity are between 10 and 40%.

17. KEY WORDS AND DOCUMENT ANALYSIS

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ABSTRACT

This project was undertaken to evaluate the development status of advanced coal combustion technologies and to prepare performance and economic models for their application to electric utility plants. The technologies addressed were atmospheric fluidized bed combustion, pressurized fluidized bed combustion, and integrated gasification combined cycle. The development status was also reviewed for pulverized coal-fired boilers incorporating supercritical steam cycles. Although the advanced combustion technologies are attractive due to decreased SO_2 and NO_x emission characteristics and potentially higher generating efficiencies, full commercial readiness appears infeasible before the mid to late 1990s. Capital cost estimates for a new plant at 500-MW ranged from $1,250 to $1,910 per kilowatt (in 1988 dollars) for the advanced technologies and $1,380 to $1,810 per kilowatt for the conventional systems. Capital cost estimates for pressurized fluidized bed (turbocharged cycle) and conventional plants with add-on SO_2 and NO_x controls were within 4 percent of the median cost ($1,580/kW) of all of the technologies evaluated. Atmospheric fluidized bed costs averaged 12 percent less than this median while integrated gasification combined cycle costs averaged 11 percent above the median. Potential capital cost savings for repowering an existing plant versus constructing a new facility at the same final capacity are between 10 and 40 percent.

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LIST OF ABBREVIATIONS AND SYMBOLS

ABBREVIATIONS

acfm	--	actual cubic feet per minute (1 acfm = 0.028 m^3/min)
AUSM	--	Advanced Utility Simulation Model
AFBC	--	atmospheric fluidized bed combustion
Btu	--	British thermal units (1 Btu = 1.055 kJ)
EHO	--	engineering and home office fees
EPRI	--	Electric Power Research Institute
ESP	--	electrostatic precipitator
FBC	--	fluidized bed combustion
FGD	--	flue gas desulfurization
gpm	--	gallons per minute (1 gpm = 0.0038 m^3/min)
HTHP	--	high temperature and high pressure
IAPCS	--	Integrated Air Pollution Control System
ID fan	--	induced draft fan
IGCC	--	integrated gasification combined cycle
kW	--	kilowatts (1 kW = 1 kJ/s)
lb	--	pounds (1 lb = 0.454 kg)
kW-h	--	kilowatt hours (1 kW-h = 3.6 MJ)
L/G	--	liquid-to-gas ratio
MCR	--	maximum continuous rating
MW	--	megawatts (1 MW = 1000 kJ/s)
MWe	--	megawatts electric
MW_t	--	megawatts thermal
NSPS	--	New Source Performance Standard

LIST OF ABBREVIATIONS AND SYMBOLS
(Continued)

ABBREVIATIONS

PC	--	pulverized coal-fired
PcsC	--	process contingency
PM	--	particulate matter
ProjC	--	project contingency
PFBC	--	pressurized fluidized bed combustion
PRCAP	--	Process capital (direct) costs
PRD	--	preproduction costs
psi	--	pounds per square inch (1 psi = 0.068 atmospheres)
SCR	--	selective catalytic reduction
scfm	--	standard cubic feet per minute (1 scfm = 0.028 m^3/min)
TPC	--	total plant costs
U. S. DOE	--	United States Department of Energy

SYMBOLS

CaO	--	calcium oxide
$CaCO_3$	--	calcium carbonate
$CaSO_4$	--	calcium sulfate
Ca/S	--	calcium-to-sulfur mole (ratio)
N_2O	--	nitrous oxide
NO	--	nitrogen oxide
NO_x	--	nitrogen oxides
SO_2	--	sulfur dioxide

SECTION 1

INTRODUCTION AND SUMMARY

1.1 INTRODUCTION

This report documents performance and economic models developed for advanced coal combustion and conventional power generation technologies. The models developed in this report are simplified to minimize computational time and data requirements, but they incorporate important parameters that have significant impacts on performance and costs. The models are based on recent information available from Electric Power Research Institute (EPRI) reports and other published sources. The sources of information are referenced in this report.

The advanced power generation technologies covered in this report include: atmospheric fluidized bed combustion (AFBC, Section 2), pressurized fluidized bed combustion (PFBC, Section 3), and integrated gasification combined cycle (IGCC, Section 4). These technologies incorporate processes for removal of sulfur dioxide (SO_2), nitrogen oxides (NO_x), and particulate matter (PM) in their designs.

The conventional power generation technology covered is pulverized coal-fired plants (Section 5). The economic model developed in this study for pulverized coal-fired (PC) plants excludes the costs for flue gas desulfurization (FGD) and selective catalytic reduction (SCR) NO_x controls. The costs for these control systems were drawn from other models and were combined with the PC base plant costs. The user has the option to include or exclude the cost of a baghouse for PM control. As such, the user of the model can add the desired control options to the uncontrolled PC boiler and compare the cost of a PC plant with those estimated for the advanced technologies.

Table 1-1 provides general descriptions of the technologies covered in this study. Also listed are the primary information sources for process descriptions, operating performance, and economic data. The FGD configuration described has been considered as the EPA "base case" system. The SCR system is a hot-side, high-ash configuration to be located upstream of the air

TABLE 1-1. GENERAL DESCRIPTIONS OF THE TECHNOLOGIES CONSIDERED

Technology	Process Description	References: Description & Performance	References: Economic Data
FGD	Forced oxidation wet limestone with in-line steam reheat; spray tower type absorber modules with liquid to gas ratio of 80 gpm/1000 acfm; three modules are assumed for 200 and 500 MWe plants and four modules are assumed for 1000 MWe plants; sludge disposal is by landfill.	1	2,3
SCR	Hot-side location between economizer and air heater; Economizer bypass to maintain nominal catalyst temperature of 700°F at low load operation; Air heater modifications to resist impact of ammonia salt formation; vertical reactor vessels with fly ash hopper and handling systems; complete catalyst replacement at 3 year intervals.	2	2
AFBC	Bubbling bed combustor; limestone sorbent; overbed coal and limestone feed system; dry waste disposal by landfill.	3,4,5,6	3,7
PFBC	Bubbling bed combustor; Dolomite sorbent; underbed coal and sorbent feed system; Turbocharged cycle; Dry waste disposal by landfill	8,9	7,8,9
IGCC	Texaco entrained flow gasifier; Oxygen blown; Radiative and convective raw gas cooler; cold gas cleanup by Selexol acid gas removal; Claus sulfur recovery; SCOT tail gas treatment; Turbine firing temperature of 2200°F	10,11,12,13	7,12,13,14

heater. The nominal NO_x removal efficiency for this system is estimated at 75 percent. The advanced technologies -- AFBC, PFBC, and IGCC -- are near-term or first generation versions of each system. Both the AFBC and PFBC combustors are of the bubbling bed design. The PFBC reference plant is a turbocharged design and does not include a gas turbine generator for combined cycle electricity production. The IGCC reference plant utilizes an oxygen-blown, entrained-flow gasifier with cold gas cleanup prior to combustion turbine firing.

Although both conventional and advanced technologies have been evaluated in this study, direct economic comparisons are difficult for several reasons. Firstly, a technology such as FGD is mature and fully commercialized with many vendors competing in the marketplace. Therefore, cost estimates for FGD systems are developed from a data base of actual installation and operating costs. In contrast, the advanced technologies, particularly PFBC and IGCC, are in relatively early stages of development with commercial service unlikely before the mid to late 1990s. Cost estimates made during such early stages of development can significantly underpredict the actual costs of the first commercial version of a new technology. However, costs for subsequent commercial installations often decline as improved versions of the technology are built. Secondly, the cost effectiveness (cost to remove a given quantity of pollutant) is much easier to define for an add-on control such as FGD or SCR than for integrated technologies. To varying degrees, processes for removal of SO_2, NO_x, and PM are intrinsic to the basic designs of the advanced generating technologies. Finally, the advanced technologies may offer significant capacity increases and increased fuels flexibility when retrofitting or repowering existing plants. These features also need to be considered when drawing comparisons between technologies.

For each technology covered in this report, the following information is presented:

- description of the technology,
- reference plant description defining the coverage of process equipment included in the cost models,
- system performance models relating to emission reduction,
- capital cost model,

- operating and maintenance cost model, and
- key technical issues - dealing with technical factors that will affect further development and use of the technology.

1.2 SUMMARY AND RESULTS

Algorithms for the performance and economic models are presented in each of the report sections corresponding to the technologies evaluated. These algorithms have also been incorporated in Version 4.0 of the Integrated Air Pollution Control System (IAPCS) cost model.[2] Version 4.0 of IAPCS was used to generate the cost presented herein.

Example cost estimates are presented for newly constructed AFBC, PFBC, and IGCC plants as well as for new pulverized coal-fired plants equipped with wet limestone FGD and SCR systems. These capital and operating cost estimates are based on early or near-term versions of the advanced technologies. As the advanced technologies are developed further, revised capital cost estimates may either increase or decrease. Revised annual operating cost estimates, however are likely to move downwards as more efficient versions of the technologies are evaluated. Cost estimates are also presented for repowering of existing pulverized coal-fired plants with the advanced coal combustion technologies. To provide comparisons for the repowering approaches, cost estimates are presented for life extension and retrofit of the existing PC plants. Equipment refurbishment and upgrade costs associated with a 20-year life extension were combined with retrofit FGD and SCR costs for the existing plants.

1.2.1 Status of Advanced Coal Combustion Technologies

The advanced coal combustion technologies have received considerable attention because of decreased SO_2 and NO_x emissions characteristics, potential improvements in overall generating efficiency, and the ability to increase generating capacity when repowering existing stations. Table 1-2 compares net plant generating efficiencies which are available currently or on

a near-term basis against projected efficiency for mature or second generation configurations for the technologies considered. The current or near-term achievable efficiencies for the advanced technologies show no improvement over conventional PC with FGD systems. However, the projected efficiencies for mature (or second generation) versions of IGCC and PFBC are about 14 to 29 percent higher, respectively, than that achievable for the conventional system. Supercritical PC plants operate at steam cycle pressures above the critical pressure of water, but utilize essentially conventional (low NO_x) pulverized coal firing systems and FGD scrubbers. Supercritical plants have been available in the U.S. since the late 1950s.[15] However, due to materials related reliability problems in the boiler superheater and reheater surfaces, main steam lines, and high pressure turbine valves, nozzle chambers, and casings, supercritical plants have not been fully commercialized.[16] Near term net plant generating efficiencies of about 39 percent are based on the performance levels of the pre-NSPS Philadelphia Electric Company Eddystone No. 1 (Eddystone, Pennsylvania) and the Ohio Power Company Philo No. 6 (near Zanesville, Ohio) plants in the late 1950s and early 1960s.[15] Proposed supercritical plant designs[17] have efficiencies of about 41 percent, or about 17 to 24 percent higher than current subcritical designs.

TABLE 1-2. NET PLANT GENERATING EFFICIENCIES[a]

Technology	Current or Near-Term Efficiency (%)	Mature Technology Projected Efficiency (%)
Conventional PC/FGD	33-35	35
Supercritical PC/FGD	39	41
AFBC	34	35
PFBC (Turbocharged Cycle)	33	35
PFBC (Combined Cycle)	34	45
IGCC	34	40

[a]Net plant efficiency is the inverse of net plant heat rate times the conversion factor, 3413 Btu/kWh.

The advanced coal combustion technologies may also be used to repower existing steam electric plants. Repowering consists of a substantial modification or replacement of the existing boiler. Where economically feasible, the existing steam turbine generator and the remaining balance-of-plant equipment is reused. In the combined cycle configurations of PFBC and IGCC, additional electricity is produced by a new gas turbine generator. Repowering can not only provide a plant service life extension, but can also result in significant capacity increases without the need to develop new greenfield power plant sites.

At their current status of development, however, the advanced technologies may not be fully available for utility application until the mid to late 1990s. Significant technical issues must be overcome before these technologies will fully penetrate the utility market. These issues include materials limitations, process control, load following ability, hot gas cleanup, and secondary environmental impacts. Table 1-3 lists some of the major issues and provides time frame estimates of commercial availability for the technologies. The development issues and availability status of the advanced technologies are more fully discussed in Sections 2, 3, 4, and 5.

1.2.2 General Costing Assumptions

The specific assumptions used to estimate cost in both constant and current 1988 dollars are shown in Table 1-4. The EPRI general cost procedures were used to incorporate inflation, cost of capital, and levelization of future expenses. Cost adjustments to account for the technical issues discussed above affecting further development and future use of each technology are not presented in this cost analysis.

Process contingencies are used to address the uncertainty associated with developing technologies while project contingencies reflect the level of detail in estimating overall project cost. In developing detailed cost estimates for a technology, process contingencies are determined for each process area or section according to its development and commercialization status. For example, in an IGCC plant, the steam turbine generator section

TABLE 1-3. DEVELOPMENT AND AVAILABILITY STATUS OF ADVANCED COAL COMBUSTION TECHNOLOGIES

Technology	Key Technical Issues	Estimated Commercial Availability	
		First Generation	Fully Mature Second Generation
Supercritical PC/FGD	Superheater materials limitations	1962	1993
	Main steam line creep and expansion failure		
	Steam turbine erosion, creep, and expansion failure		
	Load following and control		
	Plant trip critical pressure letdown		
AFBC (Bubbling Bed)	Modest combustion efficiency	1989	1995
	Poor limestone utilization		
	Part-load efficiency and emissions control		
	Solid waste production		
	Potential N_2O emissions		
PFBC	Erosion/corrosion of in-bed heat transfer surfaces	1995	2000
	Erosion/corrosion of gas turbine blades		
	Load following and control		
	Solid waste production		
	Alkali and halogen emissions		
	Potential N_2O emissions		
IGCC	Process heat recovery	1988	1996
	Solid waste production from in-situ sulfur capture		
	High temperature S, N, PM, and alkali removal		
	High temperature gas turbine development		

TABLE 1-4. ECONOMIC ASSUMPTIONS USED IN COST ANALYSIS

CURRENT DOLLARS (1988)			
Weighted Cost of Capital		12.5%	
Inflation Rate		6.0%	
Carrying Charge Factor			
Economic Life:	20 years	0.189	(Retrofit, Repowered Plants)
	30 years	0.175	(New Plants)
Levelization Factor			
Economic Life:	20 years	1.57	
	30 years	1.75	
CONSTANT DOLLARS (1988)			
Weighted Cost of Capital		6.1%	
Inflation Rate		0%	
Carrying Charge Factor			
Economic Life:	20 years	0.123	(Retrofit, Repowered Plant)
	30 years	0.105	(New Plants)
Levelization Factor			
Economic Life:	20 years	1.00	
	30 years	1.00	

INDIRECT COST	CASE 1	CASE 2
General Facilities	10%	10%
Engineering and Home Office Fees	10%	10%
Process Contingencies	10%	0%
Project Contingencies	30%	15%

will typically be assigned a process contingency at or near zero. However, given that hot gas cleanup systems have generally not progressed from pilot demonstration scale to full commercial status, a process contingency within the range of 20 to 35 percent is more appropriate for this section. A weighted average of the process area contingencies can then be expressed for the technology.

A range of contingency values was used in the present costing analysis to represent optimistic as well as conservative estimates. Weighted average contingencies are shown in the analysis for each technology rather than individual process area contingencies. Contingencies of 0 percent for process and 15 percent for project are appropriate for mature processes and detailed project cost estimates. Contingencies of 30 percent for project and 10 percent for process give more conservative cost estimates corresponding to increased process uncertainties and reduced detailed in project cost estimating. In other work,[1] EPA has used a 15 percent contingency factor for the FGD base case estimate. Contingencies inherent to the example cost estimates contained in the 1986 Technical Assessment Guide (TAG)[7] are on the order of 18 to 20 percent for AFBC and IGCC and are greater than 30 percent for PFBC. As more experience is gained in some of the process areas and with improved project costing, these factors may be refined. Contingency factors developed by EPRI following detailed engineering and risk analyses currently fall within the range of 15 to 20 percent for AFBC, IGCC, and PFBC (M. Gluckman, Electric Power Research Institute, personal communication, December 12, 1989). Each of the technologies in the present study has been evaluated using contingency values representing both optimistic and conservative premises.

Table 1-5 presents coal characteristics for a midwestern bituminous coal used in the cost analysis. To be consistent with an EPRI study of fluisized bed combustion,[3] a coal cost of 2.00 $/10^6$ Btu was used for a hypothetical Illinois coal. It should be noted that actual fuel costs may differ from this assumed price, thus affecting the relative economic ranking between the technologies. Higher fuel costs will have a greater impact on total annual operating costs for low efficiency technologies as compared to technologies with high generating efficiencies. Table 1-6 presents the unit costs for

TABLE 1-5. COAL CHARACTERISTICS AND PRICES USED IN THE COST ANALYSIS[7]

Coal Characteristics	Midwestern Bituminous Coal
Sulfur Content, %	2.0 - 4.0
Ash Content, %	16
Carbon Content, %	57.6
Heating Value, Btu/lb	10,100
Net Heat Rate for Conventional Plants with FGD, Btu/kW-h	10,060
Net Heat Rate for Conventional Plants with FGD and SCR, Btu/kW-h[a]	10,160
Net Heat Rate for AFBC, Btu/kW-h	10,000
Net Heat Rate for PFBC, Btu/kW-h	9,703
Net Heat Rate for IGCC, Btu/kW-h	9,280
Coal Price, $/ton	40.0
$/10^6 Btu	2.00

[a]Assumes that SCR will increase the net heat rate by 1 percent over that of conventional plants with FGD.

TABLE 1-6. UNIT COSTS FOR ESTIMATING OPERATING AND MAINTENANCE COSTS (1988 DOLLARS)

Item	Units	Value
Operating Labor	$/man-hour	21.40
Water	$/1000 gal	0.65
Limestone and Dolomite	$/ton	16.30
Waste Disposal	$/ton	10
Sulfur	$/ton	65
Ammonia	$/ton	150
SCR Catalyst	$/ton	20,300
Steam	$/10^6 Btu	7.00

estimating operating and maintenance costs. Annual costs were estimated using a capacity factor of 0.65. The costs are reported for the technologies firing 2 or 4 percent sulfur coal.

Table 1-7 presents the design specifications for acid gas control used to estimate the costs for each technology. For SO_2 control, 90 percent removal is assumed for AFBC, PFBC, and FGD; 95 percent removal is assumed for IGCC. For NO_x control, a NO_x emission limit of 0.6 $lb/10^6$ Btu (260 ng/J), corresponding to the current New Source Performance Standard (NSPS) level, is assumed for AFBC plants and PC plants without SCR. For PC plants with SCR, the design NO_x removal efficiency is assumed to be 80 percent corresponding to an emission level of 0.12 $lb/10^6$ Btu (52 ng/J). Measured NO_x emissions from a large demonstration PFBC boiler ranged from 0.15 to 0.5 $lb/10^6$ Btu[18]. In the present study, a NO_x emission level of 0.2 $lb/10^6$ Btu is assumed based on conceptual bubbling bed PFBC performance levels.[8]. The NO_x emission level of 0.27 $lb/10^6$ Btu (117 ng/J) for IGCC is based on the use of wet injection in the gas turbine for complying with the gas turbine NSPS limit (75 ppm at 15 percent dry O_2).[12]

1.2.3 New Plant Costs

Tables 1-8 and 1-9 present the capital and annual costs of new AFBC, PFBC, and IGCC plants firing 2 and 4 percent sulfur bituminous coals, respectively. The costs are reported for 200-, 500-, and 1000-MW plant sizes.

For comparison with the advanced coal combustion costs, costs are also reported in Tables 1-8 and 1-9 for newly constructed pulverized coal-fired plants with wet limestone FGD and SCR systems. Costs for new wet limestone FGD and SCR systems were estimated from the Integrated Air Pollution Control System (IAPCS) model.[2] The SCR costs are based on a 3-year catalyst life. Tables 1-8 and 1-9 include the operating costs of the conventional plants with baghouses for PM control. The annual costs of FGD and FGD plus SCR were added to the new plant costs so that the annual conventional plant cost can be compared directly with those for the advanced coal combustion technologies.

TABLE 1-7. DESIGN SPECIFICATIONS FOR ACID GAS CONTROL

Design Specifications	Technologies Evaluated[a]				
	AFBC	PFBC	IGCC	PC/FGD/SCR	PC/FGD
SO_2 CONTROL:					
- SO_2 Removal percent	90	90	95	90	90
- SO_2 Control Method	in-situ	in-situ	Selexol process	wet limestone FGD	wet limestone FGD
- Sorbent Type	limestone	dolomite	Selexol	limestone	limestone
- Ca/S Ratio	3.1	1.5	NA	NA	NA
- L/G Ratio	NA	NA	NA	80	80
- Stoichiometric Ratio	NA	NA	NA	1.15	1.15
- Retrofit Factor[b]	NA	NA	NA	1.5	1.5
NO_x CONTROL:					
- NO_x Emissions, lb/10^6Btu	0.6	0.2	0.27	0.12	0.6
- NO_x Control Method	in-situ	in-situ	wet injection	SCR	NA
- NH_3:NO_x Ratio	NA	NA	NA	0.82	NA
- SCR Space Velocity, 1/hr	NA	NA	NA	2,500	NA
- Number of SCR Catalyst layers	NA	NA	NA	3	NA
- SCR Catalyst Replacement, years	NA	NA	NA	2	NA
- Retrofit Factor[b]	NA	NA	NA	1.34	NA

[a]AFBC = atmospheric fluidized bed combustion.
PFBC = pressurized fluidized bed combustion.
IGCC = integrated gasification combined cycle.
PC = pulverized coal-fired plant.
FGD = flue gas desulfurization.
SCR = selective catalytic reduction.

[b]For estimating new plant costs, retrofit factor is 1.0. For estimating repowered AFBC, PFBC, and IGCC costs, costs for new plant of each technology were adjusted using the reuse and refurbishment factor approach discussed in this report.

TABLE 1-8. NEW PLANT COST ESTIMATES FOR THE ADVANCED COAL COMBUSTION TECHNOLOGIES FIRING A 2 PERCENT SULFUR BITUMINOUS COAL[a]

Technology[b]	Net Plant Capacity MW	Capital Costs $/kW	Annual Costs, mills/kW-h Constant $	Annual Costs, mills/kW-h Current $	Unit Cost per Ton of Acid Gas Removed, $/ton[c] Constant $	Unit Cost per Ton of Acid Gas Removed, $/ton[c] Current $
Case 1, Project Contingencies=15%, Process Contingencies=0%						
AFBC	200	1530	60	103	270	460
	500	1250	52	90	210	360
	1000	1080	48	82	180	300
PFBC	200	1800	68	116	580	1000
	500	1440	58	99	420	720
	1000	1220	52	89	320	560
IGCC	200	1890	65	112	470	800
	500	1540	55	95	310	540
	1000	1330	50	85	230	390
PC/FGD/SCR	200	1840	70	120	680	1150
	500	1490	61	104	550	940
	1000	1310	56	96	510	870
PC/FGD	200	1720	65	112	570	980
	500	1390	57	97	450	760
	1000	1122	52	89	400	680
Case 2, Project Contingencies=30%, Process Contingencies=10%						
AFBC	200	1860	67	115	310	530
	500	1510	58	99	240	410
	1000	1300	53	91	190	330
PFBC	200	2190	76	130	670	1150
	500	1750	64	110	490	830
	1000	1470	58	98	370	630
IGCC	200	2300	74	127	580	990
	500	1880	63	107	400	680
	1000	1610	56	96	300	520
PC/FGD/SCR	200	2150	75	128	750	1280
	500	1750	65	110	600	1030
	1000	1540	59	101	550	940
PC/FGD	200	2070	73	125	640	1100
	500	1670	63	107	490	840
	1000	1460	57	98	440	750

[a]Costs are in 1988 dollars.

[b]AFBC = atmospheric fluidized bed combustion.
PFBC = pressurized fluidized bed combustion.
IGCC = integrated gasification combined cycle.
PC/FGD/SCR = pulverized-coal plant with wet limestone flue gas desulfurization and selective catalytic reduction.
PC/FGD = PC plant with wet limestone FGD.

[c]Tons of acid gas removed = sum of the tons of SO_2 and NO_x removed.
Hypothetical acid gas control cost = annual cost for advanced technologies - annual cost for PC plant without FGD and SCR.

TABLE 1-9. NEW PLANT COST ESTIMATES FOR THE ADVANCED COAL COMBUSTION TECHNOLOGIES FIRING A 4 PERCENT SULFUR BITUMINOUS COAL[a]

Technology[b]	Net Plant Capacity MW	Capital Costs $/kW	Annual Costs, mills/kW-h Constant $	Annual Costs, mills/kW-h Current $	Unit Cost per Ton of Acid Gas Removed, $/ton[c] Constant $	Unit Cost per Ton of Acid Gas Removed, $/ton[c] Current $
Case 1, Project Contingencies=15%, Process Contingencies=0%						
AFBC	200	1550	63	108	220	380
	500	1260	55	95	190	330
	1000	1090	51	87	170	300
PFBC	200	1810	70	120	380	660
	500	1440	60	103	300	510
	1000	1220	54	93	250	420
IGCC	200	1930	65	112	260	440
	500	1570	55	95	170	290
	1000	1350	50	85	120	210
PC/FGD/SCR	200	1810	69	117	420	580
	500	1480	60	102	350	470
	1000	1300	55	95	320	420
PC/FGD	200	1690	64	110	340	580
	500	1380	56	96	280	470
	1000	1200	52	88	240	420
Case 2, Project Contingencies=30%, Process Contingencies=10%						
AFBC	200	1870	70	120	240	420
	500	1520	61	104	210	350
	1000	1310	56	96	180	310
PFBC	200	2200	78	134	430	740
	500	1750	67	114	330	570
	1000	1480	60	102	270	460
IGCC	200	2340	74	127	320	550
	500	1910	63	107	220	380
	1000	1640	56	96	160	280
PC/FGD/SCR	200	2230	80	136	460	790
	500	1810	69	113	380	650
	1000	1590	63	108	350	600
PC/FGD	200	2090	75	128	370	640
	500	1690	64	110	300	510
	1000	1480	59	101	270	460

[a]Costs are in 1988 dollars.

[b]AFBC = atmospheric fluidized bed combustion.
PFBC = pressurized fluidized bed combustion.
IGCC = integrated gasification combined cycle.
PC/FGD/SCR = pulverized-coal plant with wet limestone flue gas desulfurization and selective catalytic reduction.
PC/FGD = PC plant with wet limestone FGD.

[c]Tons of acid gas removed = sum of the tons of SO_2 and NO, removed.
Hypothetical acid gas control cost = annual cost for advanced technologies - annual cost for PC plant without FGD and SCR.

The cost effectiveness, or unit cost per ton of acid gas removed is also presented for each technology in Tables 1-8 and 1-9. The tons of acid gas removed were determined by applying the advanced technology removal efficiencies to the emissions of NO_x and SO_2 from an uncontrolled PC plant at given plant sizes. For the integrated technologies, FBC and IGCC, it is difficult to separate the costs associated with acid gas control from the rest of the costs attributed to electrical generation. Therefore, to estimate the effective costs for acid gas control for the new plants, the annual costs (including annualized capital costs) for the conventional PC plant exclusive of FGD and SCR costs were subtracted from the annual costs (including annualized capital costs) for a new AFBC, PFBC, and IGCC plant at the same final net generating capacity. This is represented by the following equation:

$$\left(\begin{array}{l}\text{Effective Acid Gas}\\ \text{Control Costs}\end{array}\right) = \left(\begin{array}{l}\text{Annual Cost for}\\ \text{Advanced Technology}\end{array}\right) - \left(\begin{array}{l}\text{Annual cost for PC plant}\\ \text{without FGD and SCR}\end{array}\right)$$

For conventional PC boilers, the cost for acid gas control is simply the annual costs for FGD and SCR.

Figures 1-1, 1-2 and 1-3 present graphically the capital costs, annual costs, and unit cost per ton of acid gas removed from Table 1-8 as a function of plant sized for the 2 percent sulfur coal case, respectively. Only constant 1988 dollar cost estimates and 30 percent for project and 10 percent for process contingencies are presented in these figures.

1.2.4 Repowered Plant Costs

Tables 1-10 and 1-11 present the capital and annual costs of repowering AFBC, PFBC, and IGCC on existing plants firing 2 and 4 percent sulfur bituminous coals, respectively. These costs are based on repowered plant final net capacity and are reported for 200-, 500-, and 1000-MW plant sizes. Whereas plant capacity is assumed to remain unchanged for AFBC and turbocharged PFBC repowering, it is assumed that IGCC repowering results in a tripling of net plant output.[20] The capital costs are based on reuse and refurbishment factors for each technology. They do not however, include the costs for replacement power during the construction outage.

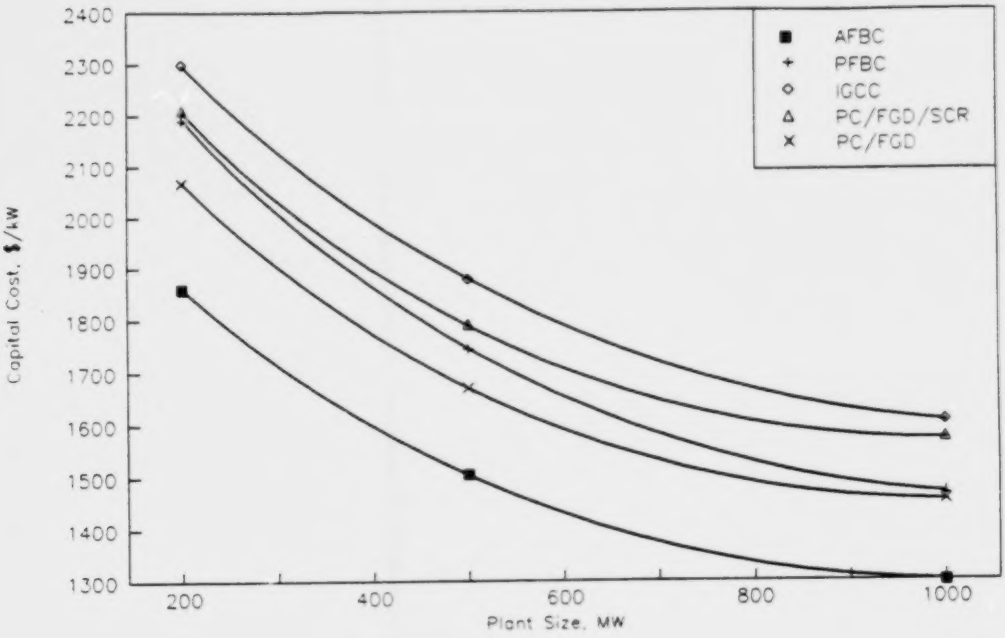


Figure 1-1. Capital costs for new plants as a function of size for plants using a 2 percent sulfur coal (in constant 1988 dollars). Contingencies of 30 percent for project and 10 percent for process are assumed.

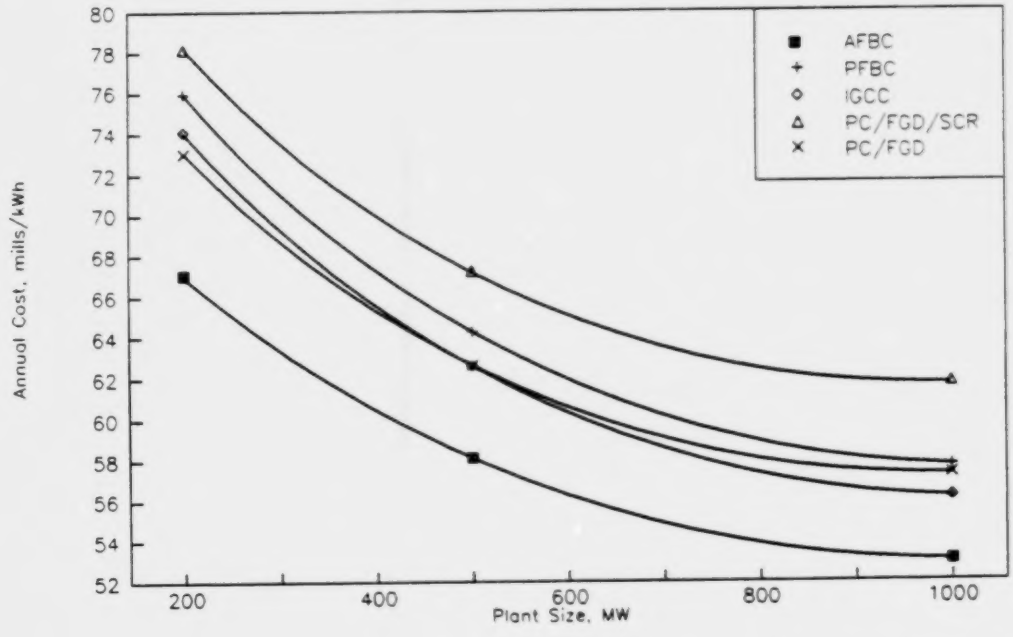


Figure 1-2. Annual costs for new plants as a function of size for plants using a 2 percent sulfur coal (in constant 1988 dollars). Contingencies of 30 percent for project and 10 percent for process are assumed.

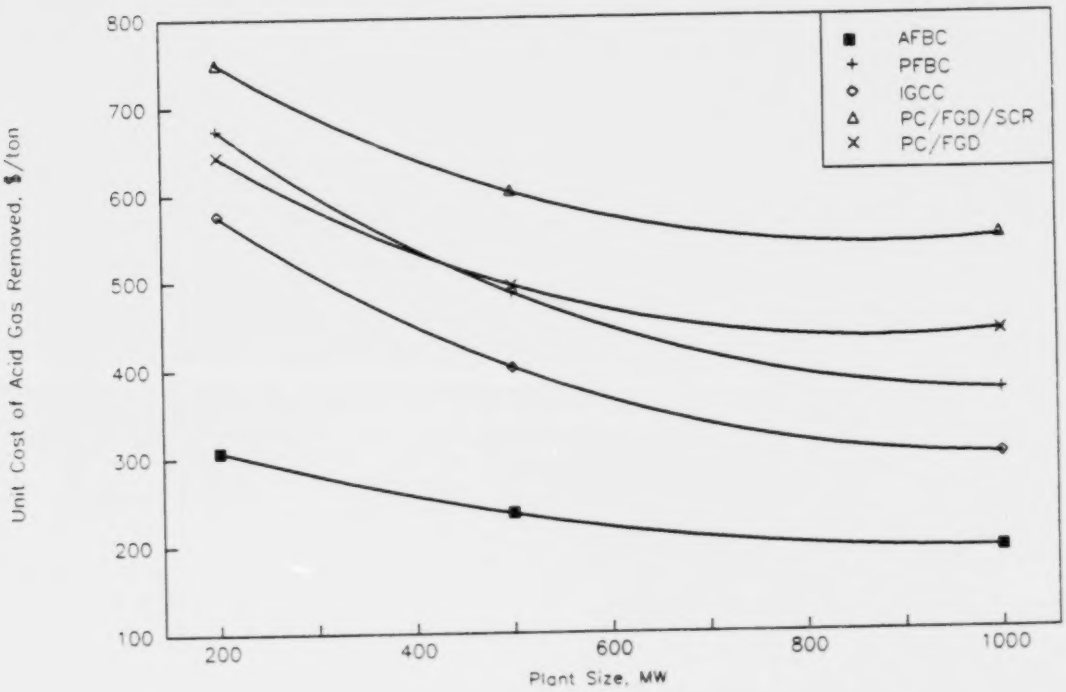


Figure 1-3. Unit cost of acid gas removed as a function of size for new plants using a 2 percent sulfur coal (in constant 1988 dollars). Contingencies of 30 percent for project and 10 percent for process are assumed.

TABLE 1-10. REPOWERED AND RETROFIT PLANT COST ESTIMATES FOR THE ADVANCED COAL COMBUSTION TECHNOLOGIES FIRING A 2 PERCENT SULFUR BITUMINOUS COAL[a]

Technology[b]	Base Net Plant Capacity MW	Final Net Plant Capacity[c] MW	Capital Costs $/kW	Annual Costs, mills/kW-h Constant $	Current $
Case 1, Project Contingencies = 15%, Process Contingencies = 0%					
AFBC	200	200	790	46	72
	500	500	660	42	65
	1000	1000	590	39	61
PFBC	200	200	1190	58	90
	500	500	950	50	78
	1000	1000	810	45	71
IGCC	70	200	1710	67	104
	170	500	1400	56	88
	330	1000	1210	51	79
PC/FGD/SCR	208	200	630	45	70
	520	500	540	40	63
	1040	1000	510	39	61
PC/FGD	206	200	490	40	62
	515	500	420	36	56
	1030	1000	400	34	53
Case 2, Project Contingencies = 30%, Process Contingencies = 10%					
AFBC	200	200	950	50	78
	500	500	800	45	70
	1000	1000	710	42	66
PFBC	200	200	1440	64	100
	500	500	1150	55	85
	1000	1000	970	49	77
IGCC	70	200	2080	76	118
	170	500	1700	64	100
	330	1000	1470	57	89
PC/FGD/SCR	208	200	700	47	73
	520	500	570	42	66
	1040	1000	550	34	53
PC/FGD	206	200	540	41	64
	515	500	450	37	57
	1030	1000	420	35	55

[a]Costs are in 1988 dollars

[b]AFBC = atmospheric fluidized bed combustion.
PFBC = pressurized fluidized bed combustion.
IGCC = integrated gasification combined cycle.
PC/FGD/SCR = pulverized-coal plant with wet limestone flue gas desulfurization and selective catalytic reduction.
PC/FGD = PC plant with wet limestone FGD.

[c]Station generating capacity is unchanged for simple cycle AFBC and PFBC repowering. Repowering with combined cycle IGCC results in an approximate tripling of capacity. Retrofit with FGD and SCR results in a capacity derate.

TABLE 1-11. REPOWERED AND RETROFIT PLANT COST ESTIMATES FOR THE ADVANCED COAL COMBUSTION TECHNOLOGIES FIRING A 4 PERCENT SULFUR BITUMINOUS COAL[a]

Technology[b]	Base Net Plant Capacity MW	Final Net Plant Capacity[c] MW	Capital Costs $/kW	Annual Costs, mills/kW-h Constant $	Current $
		Case 1, Project Contingencies = 15%, Process Contingencies = 0%			
AFBC	200	200	800	49	76
	500	500	670	44	69
	1000	1000	600	42	65
PFBC	200	200	1200	60	94
	500	500	960	52	81
	1000	1000	810	48	74
IGCC	70	200	1750	67	104
	170	500	1420	56	88
	330	1000	1230	51	79
PC/FGD/SCR	208	200	690	47	73
	520	500	590	42	66
	1040	1000	550	40	63
PC/FGD	206	200	550	41	64
	515	500	470	37	58
	1030	1000	430	36	56
		Case 2, Project Contingencies = 30%, Process Contingencies = 10%			
AFBC	200	200	960	53	83
	500	500	810	48	75
	1000	1000	720	45	70
PFBC	200	200	1440	66	103
	500	500	1150	57	89
	1000	1000	980	52	81
IGCC	70	200	2120	76	119
	170	500	1730	64	100
	330	1000	1490	57	89
PC/FGD/SCR	208	200	770	49	76
	520	500	640	44	68
	1040	1000	590	42	65
PC/FGD	206	200	610	43	67
	515	500	510	38	60
	1030	1000	470	37	57

[a]Costs are in 1988 dollars.

[b]AFBC = atmospheric fluidized bed combustion.
PFBC = pressurized fluidized bed combustion.
IGCC = integrated gasification combined cycle.
PC/FGD/SCR = pulverized-coal plant with wet limestone flue gas desulfurization and selective catalytic reduction.
PC/FGD = PC plant with wet limestone FGD.

[c]Station generating capacity is unchanged for simple cycle AFBC and PFBC repowering. Repowering with combined cycle IGCC results in an approximate tripling of capacity. Retrofit with FGD and SCR results in a capacity derate.

For comparison with the costs of repowering, costs are also reported in Tables 1-10 and 1-11 for life extension of an existing plant for 20 years of additional operation and retrofit of add-on wet limestone FGD and SCR systems and fabric filter PM control. Retrofit of FGD and SCR systems results in decreased operating efficiency which may derate the plant. In this analysis, it is assumed that an FGD retrofit gives a 3 percent capacity derate and the addition of an SCR system gives another 1 percent derate.[7] The capital cost of life extension is $214/kW, which is the average cost reported by EPRI without considering downtime costs.[19] Costs for wet limestone FGD systems were estimated from the IAPCS model. Capital costs for the FGD and SCR systems are based on a retrofit factor of 1.5 and 1.34, respectively, as shown in Table 1-7.

1.3 CONCLUSIONS

The advanced coal combustion technologies are of interest due to decreased SO_2 and NO_x emissions characteristics, potential improvements in overall generating efficiency, and the ability to increase generating capacity when repowering existing stations. Existing NSPS emission limits as well as proposed acid rain regulations are expected to be met by most AFBC, PFBC, and IGCC configurations. In addition, prevention of significant deterioration (PSD) requirements are expected to be satisfied under repowering scenarios. Projected net plant generating efficiencies for mature technology combined cycle configurations (IGCC and PFBC) are about 14 to 29 percent higher than conventional PC boilers equipped with FGD scrubbers. Advanced supercritical PC/FGD designs have generating efficiencies comparable to mature IGCC plants but still about 8 to 9 percent below second generation PFBC designs.

At their current status of development, however, the advanced technologies may not be fully available for utility application until the mid to late 1990s. Significant technical issues must be overcome before these technologies will fully penetrate the utility market. These issues include materials limitations, process control, load following ability, hot gas cleanup, and secondary environmental impacts.

Capital cost estimates for a new plant at 500 MW ranged from $1,250 to $1,910 per kilowatt for the advanced technologies and from $1,380 to $1,810 per kilowatt for the conventional systems. Capital cost estimates for pressurized fluidized bed (turbocharged cycle) and conventional plants with add-on SO_2 and NO_x controls were within 4 percent of the median cost ($1,580/kW) of all of the technologies evaluated. Atmospheric fluidized bed costs averaged 12 percent less than this median while integrated gasification combined cycle costs averaged 11 percent above the median. Potential capital cost savings for repowering an existing plant versus constructing a new facility at the same final capacity are between 10 and 40 percent.

REFERENCES

1. Skumanich, M., Etter, C., and Green, S., RCG/Hagler, Bailly, Inc. Revised Scrubber Cost Estimates. Memorandum to Tricia Pitton, U.S. Environmental Protection Agency, Office of Policy Planning and Evaluation, Washington, DC 20460. February 7, 1989.

2. Palmisano, P.J., and B. A. Laseke. User's Manual for the Integrated Air Pollution Control System Design and Cost Estimating Model, Version II, Volume I. EPA-600/8-86-031a (NTIS PB87-127767). U.S. Environmental Protection Agency, Research Triangle Park, NC. September 1986.

3. Bechtel Group Inc. Evaluation of Alternative Steam Generator Designs for Atmospheric Fluidized Bed Combustion Plants. Prepared for the Electric Power Research Institute. EPRI CS-5296. Palo Alto, CA. July 1987.

4. Aul, E.F., M. L. Owen, and A. F. Jones. Fluidized Bed Combustion: Effectiveness of an SO_2 Control Technology for Industrial Boilers, EPA-450/3-85-010 (NTIS PB85-197101). U.S. Environmental Protection Agency, Research Triangle Park, NC. September 1984.

5. Allen, R.E., Karalis, A.J., and Menaker, B.J. Comparison of a Year 2000 Atmospheric Circulating Fluidized Bed and Conventional Coal-Fired Power Plant Technical Features and Costs. Proceedings from the 1989 International Conference on Fluidized Bed Combustion. ASME. New York, NY. Vol. I, pp. 511-518, April 30 - May 3, 1989.

6. Plass, L., Beisswenger, H., Andres, R., and Lienhard, H. Operating Experience from Large Scale CFB Power Plants and Design Criteria for New CFB Power Plants. Proceedings from the 1989 International Conference on Fluidized Bed Combustion. ASME. New York, NY. Vol. II, pp. 717-728, April 30 - May 3, 1989.

7. Electric Power Research Institute. TAG-Technical Assessment Guide, Volume 1: Electricity Supply-1986. Palo Alto, CA. EPRI Report P-4463-SR, December 1986.

8. Mukherjee, D.K., et al. PFBC Turbocharged Boiler Design and Economic Study. EPRI CS-5487, Vol. 1. Electric Power Research Institute. Palo Alto, CA. November 1987.

9. Jensen, A.D., Progress Report on the Tidd PFBC Repowering Project and Comparative Economics of the P-800 PFBC System. In: Comparative Economics of Clean Coal Technologies. Pasha Publications. Arlington, VA. November 1987.

10. Clark, E., Cool Water Coal Gasification Programs, An Update. EPRI Journal. 88:45-48. September 1988.

11. Synthetic Fuels Associates, Inc. Coal Gasification Systems: A Guide to Status, Applications, and Economics. EPRI Report AP 3109. June 1983.

12. Pietruszkiewicz, J., et al. An Evaluation of Integrated-Gasification-Combined-Cycle and Pulverized Coal-Fired Steam Plants. Volume 1: Base Case Studies. EPRI AP-5950. Electric Power Research Institute. Palo Alto, CA. September 1988.

13. Pietruszkiewicz, J., et al. An Evaluation of Integrated-Gasification-Combined-Cycle and Pulverized-Coal-Fired Steam Plants. Volume 2: Sensitivity Studies and Appendices. EPRI AP-5950. Electric Power Research Institute. Palo Alto, CA. September 1988.

14. Fluor Engineers, Inc. Economic Assessment of the Impact of Plant Size on Coal Gasification-Combined-Cycle-Plants. EPRI AP-3084. Electric Power Research Institute. Palo Alto, CA. May 1983.

15. Bannister, R.L., et al. High Temperature Supercritical Steam Turbines. Mechanical Engineering, 109(2): 60-65, February 1987.

16. Bannister, R.L. and Silvestri, G.J., Jr. The Evolution of Central Station Steam Turbines. Mechanical Engineering, 111(2): 70-78, February 1989.

17. Bannister, R.L., Silvestri, G.J., Jr., and Hizume, A. Evolution and Performance of High-Temperature and Supercritical Steam Turbines. Proceedings: First International Conference on Improved Coal-Fired Power Plants. pp 3-209 to 3-229. EPRI CS-5581-SR. Palo Alto, CA. 1988.

18. Grimethorpe Pressurized Fluidized Bed Combustion Project - Overall Project Review. NCB (IEA Grimethorpe) Ltd. Barnsley, U.K. GEF/U/84/21. May 1985.

19. Electric Power Research Institute. Generic Guidelines for the Life Extension of Fossil Fuel Power Plants. EPRI CS-4778, Electric Power Research Institute, Palo Alto, CA. 1986.

20. Waddell, J.T., et al. Site-Specific Repowering Options for Fossil Energy Electric Utility Plants. Prepared for the U.S. Department of Energy. Washington, DC. Radian Final Report DCN: 89-296-002-01-02. Contract DE-AC01-88FE61666.

SECTION 2

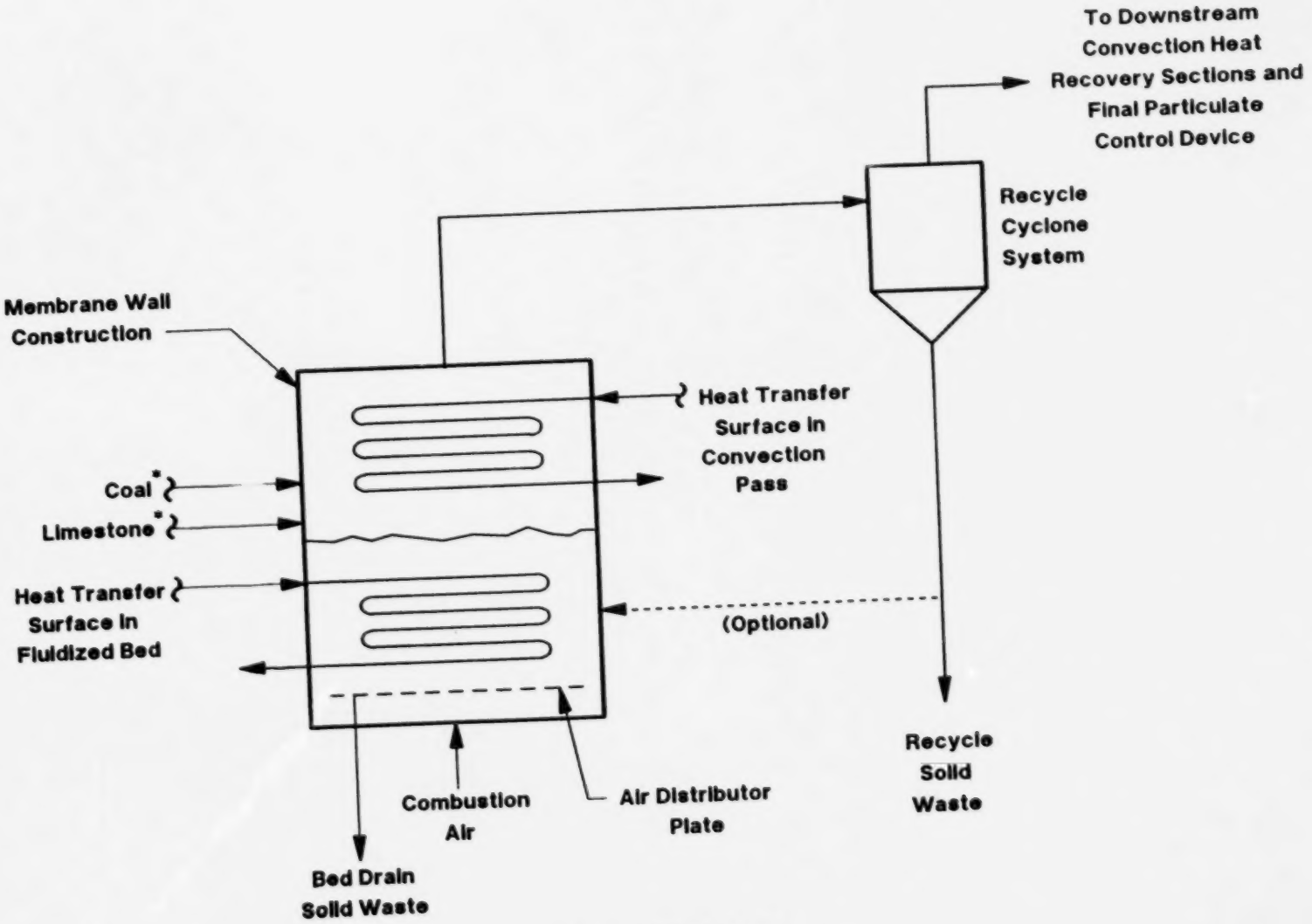
ATMOSPHERIC FLUIDIZED BED COMBUSTION (AFBC)

2.1 OVERVIEW

Atmospheric fluidized bed combustion (AFBC) is an alternative combustion technology for solid fuel-fired boilers in which fuel is burned in an air-fluidized bed of limestone or inert particles. Atmospheric fluidized bed combustion is a commercial technology for capturing SO_2 emissions from industrial boilers. AFBC technology could attain commercial status for utility boilers in the next few years if utility-scale demonstration projects currently in operation or under construction prove successful.

Figure 2-1 presents a simplified AFBC process flow diagram. Coal and limestone are fed into a bed of hot particles (1,400-1,600°F) fluidized by upflowing air. Sulfur dioxide formed during combustion reacts with the calcined limestone to form calcium sulfate, thus reducing SO_2 emissions and avoiding the need for post-combustion controls. The relatively low combustion temperature limits NO_x formation, reduces ash fusion problems, and optimizes sulfur capture. One of the major advantages of AFBC technology is its ability to burn a wide range of fuels, including those with low heat content and high sulfur.

There are two major AFBC types: the bubbling or dense bed, and the circulating or dilute bed. In the bubbling bed system, coal and limestone are continuously fed into the boiler from over the bed (denoted as overbed feed boilers) or under the bed (denoted as underbed boilers). The bed materials--consisting of unreacted, calcined, and sulfated limestone; coal; and ash--are suspended by the combustion air blowing upward through the air distributor plate. Bed material is drained from the bed to maintain the desired bed depth. Some bed material is entrained in the upflowing flue gas and escapes the combustion chamber. Approximately 80 to 90 percent of this fly ash is collected in the cyclone and is then either discarded or



* Coal and limestone may be fed above, in, or under the fluidized bed.

Figure 2-1. Conventional AFBC boiler flowsheet.

reinjected into the bed. Reinjection of ash improves combustion efficiency and limestone utilization. In general, combustion efficiency increases with longer freeboard residence times and greater ash recycle rates. (The freeboard is defined as the space between the top of the fluidized bed and the boiler outlet.) Fly ash not collected in the cyclone is removed from the flue gas by an electrostatic precipitator (ESP) or fabric filter.

The circulating fluidized bed is a more recent development in AFBC technology. The two major differences between circulating and bubbling AFBC units are the size of the limestone particles fed to the system and the velocity of the fluidizing air stream. Limestone feed to a bubbling bed is generally less than 0.1 inches in size, whereas circulating beds use much finer limestone particles, generally less than 0.01 inches. The bubbling bed also incorporates relatively low superficial air velocities, ranging from 4 to 12 ft/sec. This creates a relatively stable fluidized bed of solid particles with a well-defined upper surface. Circulating beds have superficial velocities ranging from 20 to 40 ft/sec. As a result, a physically well-defined bed is not formed; instead, solid particles (coal, limestone, ash, sulfated limestone, etc.) are entrained in the transport air/combustion gases. These solids are then separated from the combustion gases by a cyclone or other separating device and circulated back into the combustion region, along with fresh coal and limestone. A portion of the collected solids is continuously removed from the system to maintain material balances. Circulating beds are characterized by very high recirculated solids flow rates, up to three orders of magnitude higher than the combined coal/limestone feed rate.[1]

Several advantages of the circulating bed over the bubbling bed have been identified. These are:

- higher combustion efficiency, exceeding 99 percent,
- greater limestone utilization, due to high recycle of unreacted sorbent and small limestone feed size (Greater than 85 percent SO_2 removal efficiency is projected with a calcium-to-sulfur (Ca/S) ratio of about 1.5, with the potential for greater than 95 percent SO_2 removal efficiency.),

- lower NO_x emissions because of staged combustion (less than 100 ppm NO_x is projected),
- simpler coal feed design, since high velocities ensure good mixing,
- potentially fewer corrosion and erosion problems, since heat transfer surface is less likely to be located in the primary combustion zone,
- minimal excess air requirements, since the high velocities promote good mixing and combustion efficiency,
- less dependence on limestone type, since reactivity is improved with the fine particle sizes, and
- reduced solid waste rates, because of lower limestone requirements.

Potential disadvantages are:

- increased capital costs,
- greater energy losses due to high pressure drops across the systems,
- increased combustor height compared to a bubbling bed,
- uncertainty regarding the hot cyclone's ability to adequately separate solids and gases and to resist erosion and corrosion,
- greater difficulty in using existing boiler equipment in retrofit applications, and
- erosion of components subjected to impingement of high velocity particles.

2.2 AFBC REFERENCE PLANT

Since most of the cost data currently available for utility AFBC systems are based on bubbling bed technology, the cost estimates developed in this analysis apply exclusively to bubbling bed boilers. Cost algorithms for new and repowered AFBC plant overbed AFBC plants were derived from Reference 2. A "greenfield" plant is defined simply as a new plant which

provides new capacity. Costs for a repowered AFBC plant include the costs for any additional equipment and refurbishment of existing equipment necessary to convert a conventional coal-fired power plant to AFBC technology.

2.3 DETAILS OF PHYSICAL PERFORMANCE MODELS

Fuel characteristics are the main factors affecting AFBC costs. In this analysis, the primary characteristic considered is the coal sulfur content. The amount of sulfur in the coal determines the level of control necessary for compliance with the NSPS for utility boilers. The percent SO_2 reduction required is achieved by maintaining the limestone feed rate at a set (Ca/S) ratio. As the Ca/S ratio increases, the limestone feed rate, and thus sorbent cost, increases. Solid waste generation also increases with increasing Ca/S ratios.

Many other factors determine the SO_2 control performance of AFBC systems, including bed temperature, the amount of fly ash recycle, limestone particle size and reactivity, and feed mechanisms. However, it is beyond the scope of this analysis to account for all these parameters in a costing model. A simple model was developed from data prepared for EPRI to estimate the Ca/S ratio required for a specified percent SO_2 reduction[2]:

$$CAS = EFFSO_2/EFFCA \quad (1)$$

where CAS = Ca/S ratio
$EFFSO_2$ = SO_2 removal efficiency (percent)
EFFCA = calcium utilization efficiency (percent),
29.2 percent for bituminous and subbituminous coals, and;
24.3 percent for lignite coal.

The data reported show calcium utilization to be different for bituminous and subbituminous coals compared to lignite coals.[2] Calcium utilization ranged from 26.9 to 31.8 percent for bituminous and subbituminous coals. The average value of 29.2 percent is used in Equation 1 for these coal types. On the other hand, calcium utilization ranged from 23.3 to 25.2 for the lignite coals. The average value of 24.3 percent is used in Equation 1 for lignite coals. For 90 percent SO_2 removal, the CAS ratio calculated by equation 1 is 3.1 for bituminous and subbituminous coals and 3.7 for lignite coals. The limestone feed rate can then be calculated by the following equation:

$$LSFR = M_{coal} * (\%S/100) * CAS * 100.09/(32.06 * 0.9) \quad (2)$$

where LSFR = limestone feedrate (tons/hr)

%S = percent sulfur in coal

M_{coal} = coal rate (tons/hr), calculated by the following:

$$M\ coal = (MW * NHR)/(HV*2) \quad (3)$$

where NHR = AFBC net heat rate (Btu/kWh)

HV = coal higher heating value (Btu/lb)

MW = net plant megawatt rating.

This estimate is based on a limestone composition of 90 percent calcium carbonate ($CaCO_3$).

The amount of solid waste generated at an AFBC site depends on the ash content of the coal, the limestone feed rate, and the efficiency of the PM control device. The solid waste consists primarily of coal ash, limestone inerts, and calcium sulfate ($CaSO_4$), which is a reaction product of the calcined limestone and SO_2. The Equation 4 takes these components into account for new AFBC units:[3]

$$SWGR = [0.9 + 0.1 * (EFFPM/100)] * \{0.624 * LSFR + M_{coal} * [(\%S) * (EFFSO_2) * 2.5/10,000 + (\%A/100)]\} \quad (4)$$

where SWGR = solid waste generation rate (tons/hr)
EFFPM = PM control efficiency (percent)
%A = coal ash content (percent).

Equation 4 is based on limestone composed of 90 percent $CaCO_3$ and a recycle cyclone efficiency of 90 percent.

2.4 CAPITAL COST MODEL

2.4.1 Process Capital Costs

The process capital costs equations for a new AFBC facility are based on estimates from EPRI for the 200 and 1,000 MW plants firing high sulfur Midwestern (bituminous) coal and located at Kenosha, Wisconsin. The 1,000 MW plant consists of two 500 MW units. EPRI provided enough detailed costs so that the costs for one 500 MW unit could be determined from the 1,000 MW plant.[2] Table 2-1 presents the process capital cost equations in million 1987 dollars for the eight process areas. Each equation in Table 2-1 is a function of key operating and design parameters that affect costs. To estimate costs at other plant sizes, power curve fits were used to relate costs to these parameters:

$$COSTS = a\ x^b \quad (5)$$

where a = the coefficient
b = the scaling factors
x = the independent variable.

TABLE 2-1. PROCESS CAPITAL COST EQUATIONS FOR OVERBED FEED AFBC

Process Area	Equipment Included	Process Area Capital Cost Equation, 10^6 (1987$)[a]
Steam Generator	Steam generator	$C = 0.042 * (BHI)^{0.915}$
PM Controls	Baghouse, ID fan, and stack	$C = 39.593 * (BHI/HV)^{0.727}$
Waste Handling/ Disposal	Solid waste handling	$C = 0.2993 * (M_{coal} * \frac{\%A}{100} + LSFR)^{0.682}$
Steam/Boiler Feed Water/Condensate	Condenser, rotating equipment, heater and exchangers, tanks and vessels, water treatment/ chemical feed, piping, and control/instrumentation	$C = 0.059 * (BHI)^{0.824}$
Steam Turbine	Steam turbine generator	$C = 0.125 * (BHI)^{0.680}$
Coal Handling/ Preparation	Coal receiving/handling, storage, and conveying	$C = 51.051 * (BHI/HV)^{0.913}$
Limestone Handling	Limestone receiving/ handling and storage	$C = 1.273 * (LSFR)^{0.352}$
General Facilities	Concrete, architectural and structural, earthwork, piles, site preparation, HVAC[b], electrical equipment, bulk materials, switchyard, and miscellaneous	$C = 0.428 * (BHI)^{0.684}$

[a] C = Process area capital costs;
BHI = Boiler heat input, 10^6 Btu/hr calculated by: $BHI = \frac{MW * NHR}{1000}$.
M_{coal} = Coal rate, tons/hr (calculated by Equation 2).
%A = Ash content of coal, wt. %.
LSFR = Limestone feed rate, tons/hr (calculated by Equation 2).
HV = Higher heating value of coal, Btu/lb.
MW = Net plant output, megawatts.
NHR = Net heat rate, Btu/kW-h.

[b] HVAC = heating, ventilating, and air conditioning.

The costs reported to EPRI[2] were updated to average 1987 dollars (from end of 1982 dollars) using the Chemical Engineering Plant Cost Index.

The process capital cost model for repowered AFBC plants was developed using the following approach:

- For each process area, a reuse factor was assigned based on the percentage of the process area that would not be needed due to reuse of the existing plant. The assignment of the reuse factor was based on existing equipment compatibility with the new hardware as well as general condition of the existing equipment. Reuse factors are very site-specific reflecting differences in plant configuration, age, and maintenance histories. Given sufficient information on a specific plant, a capital cost estimate for repowering can be constructed for that site. Generic reuse factors are estimates of average equipment compatibility and condition for the population of existing plants suitable for repowering. Using generic reuse factors, a population-wide average capital cost estimate can be prepared for a hypothetical plant. Table 2-2 presents the generic reuse factors for the AFBC process areas. A reuse factor of 1.0 indicates that reuse of the existing equipment will provide full credit or completely offset the need for new equipment within the process area. A reuse factor of 0.0 indicates that there is no equipment suitable for reuse within the process area.

- The reuse factor for each process area was multiplied by the percentage of total process cost presented by the process area to obtain a weighted reuse credit.

- The sum of the weighted reuse credits for all process areas is the percent reduction in new AFBC plant capital cost resulting from reuse of existing plant equipment.

- It was assumed that reusable equipment will be upgraded and refurbished concurrent with the repowering program. Refurbishment costs were estimated at 20 percent of the reusable equipment value. In effect, this reduces the weighted reuse credit by 20 percent. Refurbishment costs also include the costs of demolition and removal of some existing equipment.

The following equation was developed to calculate total process capital costs for a repowered AFBC plant:

$$\text{PRCAP (Repowered AFBC)} = \text{PRCAP (New AFBC)} - (1-0.2) * [\text{C (New AFBC)} * \text{Equipment Reuse Factor}] \quad (6)$$

TABLE 2-2. EQUIPMENT REUSE FACTOR[a] AS A PERCENTAGE OF NEW AFBC PLANT COST

AFBC Process Area	Equipment Reuse Factor[b]
Steam Generator	0[c]
PM Controls	0.5[d]
Waste Handling/Disposal	0.7[e]
Steam/Boiler Feedwater/Condensate	0.8[f]
Steam Turbine	0.8[g]
Coal Handling/Preparation	0.9[h]
Limestone Handling	0.0[i]
General Facilities	0.9[j]

[a]Total plant generating capacity is assumed to remain constant after repowering.

[b]Source: Snow and Maibodi[4]; Waddell et al.[5]

[c]Assumes complete boiler replacement with a new AFBC steam generator. For a bubbling bed rebuild of existing boiler, use 0.6.

[d]If there is no existing ESP or fabric filter, use 0.

[e]Assumes 70% of the process area capital cost is offset by existing reusable ash handling and disposal equipment.

[f]Assumes the existing boiler feedwater and condensate systems are well maintained and compatible with the AFBC boiler configuration.

[g]Assumes the existing steam turbine and generator are well maintained and are substantially reused.

[h]Assumes 90% of the process area capital cost is offset by existing reusable coal receiving, storage, crushing, and delivery equipment because of insignificant capacity increases.

[i]Assumes no existing limestone handling equipment for an FGD scrubber.

[j]Assumes 90% of the process area capital cost is offset by existing reusable general facilities because of insignificant increases.

where PRCAP (Repowered AFBC) - Total process capital costs for repowered a AFBC plant (10^6 1987)

PRCAP (New AFBC) - Total process capital costs for a new AFBC plant (10^6 1987)

C (New AFBC) - Individual process area capital costs for a new AFBC plant (10^6 1987).

2.4.2 Indirect Capital Costs

The total capital requirement for the project is determined by adding indirect costs to the total process capital costs. Indirect capital costs can be calculated using the EPRI methodology.[6] Table 2-3 presents the indirect costs included in the EPRI methodology. Costs for estimating engineering and home office expenses, process contingency, and royalties were based on information from EPRI.[2] Both high and low contingency values were used to give cost ranges.

The cost equations in Table 2-1 were compared to costs for units firing other coals also prepared for EPRI.[2] A 10 percent process contingency and a 15 percent project contingency was used in the cost model for this comparison. Table 2-4 presents the results of the cost comparison for a 200 MW AFBC plant firing Illinois No. 6 coal, West Virginia bituminous coal, Wyoming subbituminous coal, North Dakota lignite, and Texas lignite. As shown, the predicted total capital requirements were, 2 to 10 percent greater than the reported costs. Therefore, the cost model is a conservative but reasonable predictor of costs for units firing a range of coals. The difference in costs may be attributable to location factors such as labor costs and labor force productivity.

TABLE 2-3. INDIRECT CAPITAL COSTS FOR AFBC SYSTEMS: EPRI METHODOLOGY[6]

(1) Process Capital Costs (PRCAP)	
(2) Total Plant Costs	
General Facilities (GF)	= 10% of PRCAP
+ Engineering and Home Office (EHO)	= 10% of PRCAP
+ Process Contingency	
Case 1 (PcsC)	= 10% of PRCAP
Case 2 (PcsC)	= 0% of PRCAP
+ Project Contingency	
Case 1 (ProjC)	= 30% of (PRCAP + GF + EHO + PcsC)
Case 2 (ProjC)	= 15% of (PRCAP + GF + EHO + PcsC)
+ PRCAP	
Total Plant Costs (TPC)	= PRCAP + GF + EHO + PcsC + ProjC
(3) Royalties	= 0.5% of steam generator process area costs
(4) Preproduction Costs	
1 month fixed operating costs	
+ 1 month variable operating costs (full capacity)	
+ 25% of full capacity fuel cost for 1 month	
+ 2% of Total Plant Investment	
Preproduction Costs (PRD)	
(5) Inventory Capital = 60 days of fuel and consumable inventory at full load + 0.5% of TPC	
(6) Interest during construction = 6.2% of TPC[a]	
(7) Land = $6,500/acre	

[a]Based on a 3-year construction period at a 6.1 percent interest rate and 0 percent inflation and assuming that uniform payments occur at the end of the year.

TABLE 2-4. COMPARISON OF THE REPORTED AND PREDICTED CAPITAL REQUIREMENTS FOR A 200 MW OVERBED FEED AFBC UNIT

Plant Location	Coal Type	Total Capital Requirement, $/kW (1988$) Reported[a]	Predicted[b]	Percent Error[c]
Kenosha, WI	Illinois No. 6 Bituminous	1,535	1,683	9.6
Bethlehem, PA	West Virginia Bituminous	1,585	1,633	3.0
Kenosha, WI	Wyoming Subbituminous	1,569	1,647	5.0
Velva, ND	North Dakota Lignite	1,647	1,682	2.1
Freeport, TX	Texas Lignite	1,571	1,689	7.5

[a]Costs reported in Reference 2 in 1982$ and updated to 1988$ for presentation here.

[b]Procedures for estimating total capital requirement are the same as those used to estimate reported capital requirement using the EPRI methodology.

[c]Percent Error = 100 * (Predicted - Reported)/Reported.

2.5 OPERATING AND MAINTENANCE COST MODEL

The direct operating costs associated with AFBC systems include costs for fuel, sorbent, labor, maintenance, water, electricity (parasitic load), and solid waste disposal. The fuel costs for AFBC systems can be estimated using the equations:

$$FC = 8.76 * CF * MW * HR * CC \quad (7)$$

where FC = annual fuel costs ($/yr)
CF = capacity factor (fraction)
CC = coal cost rate ($/$10^6$ Btu).

Sorbent costs can be determined by:

$$ALC = LSFR * CF * 8760 * LSC \quad (8)$$

where ALC = annual limestone cost ($/yr)
LSC = limestone cost rate ($/ton).

From EPRI operation labor requirements for AFBCs[2], the following model was developed for this study to estimate operating labor as a function of plant size:

$$OL = 34{,}816 * (MW)^{0.307} \quad (9)$$

where OL = annual operating labor (man-hrs/yr)

Supervisory labor requirements are estimated to be 20 percent of the operating labor costs. Direct labor costs are estimated by applying the appropriate labor cost rates and labor factors. Maintenance costs are estimated as two percent of the total plant costs based on EPRI methodology.

Water requirements for AFBC units are similar to those for equal sized boilers of different design. Based on estimated water requirements for the 200 and 1,000 MW plant[2], the following equation was developed.

$$AWC = 8.76 * CF * WTR * WCR \quad (10)$$

where AWC = annual water costs ($/yr)
WCR = water cost rate ($/1,000 gal)
WTR = water requirement (gal/hr), calculated by:
WTR = -5,335 + 601.6 MW.

Solid waste generation costs for new AFBC are estimated by Equation 11:

$$SW = 8,760 * CF * SWGR * SWC \quad (11)$$

where SW = annual solid waste disposal cost ($/yr)
SWGR = solid waste generation rate (tons/yr), from Equation 4
SWC = solid waste disposal cost rate ($/ton).

The indirect operating costs include overhead and annualized capital costs. The EPRI procedure should be used to estimate the indirect operating costs.[6]

2.6 KEY TECHNICAL ISSUES

There are currently three major (greater than 100 MW) utility boiler AFBC demonstration units in the U.S. These projects are located at Northern States Power's Black Dog plant near Nicols, Minnesota; Colorado-Ute Electric Association's Nucla Station at Nucla, Colorado; and Tennessee Valley Authority's Shawnee Station at Paducah, Kentucky. All three of these applications are EPRI sponsored demonstration programs designed to refine the technology for widespread use. The Colorado-Ute Nucla demonstration is

currently operating under U. S. DOE Clean Coal Technology Round 1 funding. The following discussion highlights the technology operating performance and efficiency expectations, and the key environmental and engineering issues.

The current demonstration programs have identified a number of important technical and economic issues which will have to be resolved before AFBC is widely used by the utility industry. It is expected that these issues can be satisfactorily addressed in the 1995-2000 time frame.

2.6.1 Operating Performance and Efficiency

Nominal net heat rates for AFBC range from 9,690 to 10,045 Btu/kWh. To achieve heat rates of 9,698 Btu/kW-h (equivalent to a net plant efficiency of 35.2 percent), conceptual designs require a superheat temperature of 1,010°F at 2,640 psig with a single reheat to 1,010°F. This design configuration is not expected to be commercialized for AFBC before the year 2000.[7]

For bubbling AFBCs, mixing limitations in the bed have the greatest negative impact on performance. System complexity and capital cost tend to limit the maximum number of feed points above or below the bed. Poor mixing in the bubbling bed allows combustibles and other reactants to escape the bed and burn or react in the freeboard space. The freeboard may also be poorly mixed resulting in increased CO emissions and carbon carryover. In addition, since freeboard combustion occurs above the heat transfer tubes, temperatures can reach beyond 2,000°F giving poor SO_2 capture performance and increased NO_x emissions. Increasing the number of feed points to overcome mixing limitations requires additional fuel injection controls such as rotary valves or lockhopper system. This will increase capital cost estimates for fuel handling equipment. These issues have been largely rectified in circulating FBCs in which gaseous-and-solid phase reactants are more vigorously mixed.

Load following and part-load efficiency (heat rate) is also an operational limitation for bubbling FBC units. Load turndown is primarily accomplished by reducing fuel/air ratio and thus decreasing bed temperature,

by reducing bed height and uncovering in-bed tubes, or by slumping portions of the bed. These approaches may have negative implications on emissions performance and boiler efficiency. Bed temperature reductions below 1,500°F causes combustion efficiency to drop with corresponding increases in CO emissions and ash fixed carbon.

Lower bed heights reduce gaseous residence time for SO_2/sorbent reaction giving increased SO_2 emissions. Reduction of bed height to uncover heat transfer tubes may result in boiler efficiency losses since the tube bank may contain evaporator, superheat, and reheat sections. In addition, since heat transfer performance depends on a complex interaction of temperature, bed density, bubble properties, particle size, and local particle velocities, optimal design of the tube bank configuration and control schemes for part-load operation is difficult to accomplish.

For single bed bubbling FBC units, load turndown to approximately 40 percent of maximum continuous rating (MCR) can generally be achieved without significant degradation in efficiency and emissions performance. Large industrial and demonstration circulating FBC plants (600,000 lb/hr or greater) have also demonstrated low load operation down to 40 percent MCR.[8] Turndown below 40 percent MCR is often accompanied by heat rate degradation of greater than 10 percent. Advanced multiple bed or multiple compartment designs have achieved turndown ratios up to 12 to 1 (approximately 8 percent MCR) at laboratory scale.[9] However, efficiency or emissions performance projections were not made at such low loads. Greater levels of overall plant turndown may be achieved by designing multiple unit installations.

2.6.2 Environmental Impacts

Low emissions of SO_2 and NO_x are potential benefits of FBC. Efficient SO_2 capture depends on good mixing and sufficient residence time in the bed. Sulfur reductions of 70 to 90 percent for atmospheric bubbling FBCs are common for industrial boiler scale applications. Utility boiler scale (greater than 100 MW) experience is currently limited to one circulating bed and one bubbling bed application using low sulfur western coals.

Emissions of N_2O are another concern facing both bubbling and circulating bed FBCs. Available data are limited; however, N_2O emissions at the same order of magnitude as NO (between 50 and 200 ppm) have been reported.[10] These data were obtained with measurement procedures designed to avoid the N_2O measurement artifact (N_2O formation in the sampling flask) recently identified.

Solid waste production and disposal is another major environmental and economic concern facing FBC technology. As higher sulfur fuels are utilized and Ca/S ratios are increased to meet emission standards, the mass of waste material produced increases significantly.[11] In addition to the volume, the spent sorbent is mixed with both the bottom ash and fly ash when it is generally more desirable to produce and collect these streams separately. This mixed waste steam contains reactive lime, reduced sulfur compounds, and carbon, as well as trace metals from the coal ash. Wetting of the material to control dusting creates corrosive calcium hydroxide and substantially increases the weight. In addition, the ash has been formed at low combustion temperatures and is not sintered as in the case of PC boiler ash. This ash is therefore more leachable than PC boiler ash creating further disposal concerns. Finally, although European researchers are investigating the use of the waste as a building material aggregate, the waste is generally of poor pozzolanic quality and highly leachable because of its carbon content and lime content (especially for boilers firing high sulfur coals). Many industrial FBCs are finding waste disposal costs to be significantly higher than projections.

2.6.3 Engineering Issues

Most of the minor mechanical problems associated with early bubbling and circulating FBC units have been solved by vendors and users. The major problem which remains is erosion of in-bed heat transfer surfaces. Both atmospheric and pressurized units sometimes exhibit the phenomenon, and reliable guidelines do not exist for designing and operating FBC's to provide adequate tube lifetimes. Average tube wastage rates are on the

order of 0.3-0.5 mm per 1,000 hour. Maximum localized wear rates of up to 1.6 mm per 1,000 hour have been reported.[12] These rates are unacceptable for typical heat transfer tube thickness of 5 mm and have serious implications for reliability. Erosion remedies employed to date primarily are in the form of sacrificial metal protectors, cladding, and refractory coating. Further investigation is underway to develop and understand tube bank configuration, tube spacings, operating regimes to alleviate this problem. A potential outcome will be the requirement for more exotic tube materials and reduced fluidization velocities (increased boiler volumes), both affecting capital costs, and more careful selection of fuels and sorbents which will impact operating costs.

Fuel and sorbent feeding as well as ash removal using rotary valves, lockhoppers, or screw pumps have reliability problems associated with plugging and erosion. Problems have been particularly difficult where feed and removal points are at high temperature and pressure. Again, costs are impacted by the requirement for more expensive materials, greater redundancy, and increased maintenance. The cost model presented in Section 2.0 does not reflect the costs for more expensive materials, greater redundancy, and increased maintenance.

REFERENCES

1. Yerushalmi, J. Circulating Fluidized Bed Boilers. IN: Proceedings of the 88th National Meeting of the AIChE (Volume 1), Philadelphia, PA. 1980, pp. 490-521.

2. Bechtel Group, Inc. Evaluation of Alternative Steam Generator Designs for Atmospheric Fluidized Bed Combustion Plants. Prepared for the Electric Power Research Institute. EPRI CS-5296. Palo Alto, CA. July 1987. pp. 4-53 to 4-111, Appendix C.

3. Aul, E.F., M.L. Owen, and A.F. Jones. Fluidized Bed Combustion: Effectiveness of an SO_2 Control Technology for Industrial Boilers. EPA-450/3-85-010 (NTIS PB85-197101). U.S. Environmental Protection Agency, Research Triangle Park, NC. September 1984. p. A-5.

4. Snow, G. C., and M. Maibodi. Feasibility Analysis of Advanced Technology Repowering of Coal-Fired Utility Plants. Paper presented at the Joint ASME/IEEE Power Generation Conference, Dallas, TX. October 22-26, 1989.

5. Waddell, J. T., et al. Site-Specific Repowering Options for Fossil Energy Electric Utility Plants. Prepared for the U.S. Department of Energy. Washington, DC. Radian Final Report DCN 89-296-002-01-02. Contract DE-AC01-88FE61666. March 1990.

6. Electric Power Research Institute. TAG-Technical Assessment Guide, Volume 1: Electricity Supply-1986. Palo Alto, CA. EPRI Report P-4463-SR. December 1986. pp. 3-1 to 3-14.

7. Allen, R.E., Karalis, A.J., and Menaker, B.J. Comparison of a Year 2000 Atmospheric Circulating Fluidized Bed and Conventional Coal-Fired Power Plant Technical Features and Costs. Proceedings from the 1989 International Conference on Fluidized Bed Combustion. ASME. New York, NY. Vol. I, pp. 511-518, April 30 - May 3, 1989.

8. Plass, L., Beisswenger, H., Andres, R., and Lienhard, H. Operating Experience from Large Scale CFB Power Plants and Design Criteria for New CFB Power Plants. Proceedings from the 1989 International Conference on Fluidized Bed Combustion. ASME. New York, NY. Vol. II, pp. 717-728, April 30 - May 3, 1989.

9. Brown, R.C., Foley, J.E., and Buttermore, W. Load Turndown in a Two-Bed Fluidized Combustor. Proceedings from the 1989 International Conference on Fluidized Bed Combustion. ASME. New York, NY. Vol. II, pp. 1169-1174, April 30 - May 3, 1989.

10. Amand, L.E., and Andersson, S. Emissions of Nitrous Oxide (N_2O) from Fluidized Bed Boilers. Proceedings from the 1989 International Conference on Fluidized Bed Combustion. ASME. New York, NY. Vol. I, pp. 49-56, April 30 - May 3, 1989.

11. Makansi, J. Users Pause, Designers Wrestle With Fluid Bed Boiler Scaleup. Power. pp. 41-42, July 1988.

12. Grimethorpe Pressurised Fluidised Bed Combustion Project - Overall Project Review. NCB (IEA Grimethorpe) Ltd., Barnsley, U.K., GEF/V/84/21, May, 1985.

SECTION 3

PRESSURIZED FLUIDIZED BED COMBUSTION (PFBC)

3.1 OVERVIEW

Pressurized fluidized bed combustion is similar to AFBC except that the bed is fluidized using pressurized combustion air. A compressor increases the combustion air pressure to between 6 and 16 atmospheres. Dolomite or limestone is normally injected in the furnace for capturing the SO_2 in the combustion gases. Steam raised in the boiler and superheater tube banks is used to drive a steam turbine generator as in a conventional utility boiler. The high-temperature, high-pressure flue gas leaving the furnace is sent to a mechanical collector for removal of the suspended particulates. The cleaned flue gas drives a gas turbine expander to generate shaft power. In the turbocharged PFBC configuration, this shaft power is used only to drive the air compressor. In the combined cycle configuration, the shaft power turns a generator for producing additional electricity, thus increasing the overall generating efficiency. Further cleanup of the flue gas may be accomplished upstream of the gas turbine with an advanced, high-temperature, high-pressure ESP or ceramic filter; or downstream of the turbine using a more conventional ESP or fabric filter.

Pressurized fluidized bed combustion is not as well developed as AFBC. Initial utility PFBC demonstration projects are not expected to be in operation until the early 1990s with the earliest commercial availability projected in the 1995-2000 time frame.[1]

3.2 PFBC REFERENCE PLANT

Costs for new and repowered PFBC units are based on data in Reference 2, 3, and 4. Because of the limited amount of information available on combined cycle PFBC costs, the economic model developed here is based on

the turbocharged cycle. A combined cycle plant will require additional capital expenditures associated with the gas turbine generator system. A "greenfield" plant is defined simply as a new plant which provides new capacity. Costs for a repowered PFBC plant includes the costs for any additional equipment or refurbishing of existing equipment necessary to convert a conventional coal-fired power plant to PFBC technology.

3.3 DETAILS OF PHYSICAL PERFORMANCE MODELS

Cost algorithms are presented in Section 3.4 and 3.5 for new PFBC units based on published data.[2-4] The primary physical characteristic considered in the capital cost algorithms is generating capacity (i.e., megawatts).

Coal sulfur content primarily affects mostly operating and maintenance costs. The amount of sulfur in the coal determines the level of control necessary for compliance with the NSPS for utility boilers. The percent SO_2 reduction required is achieved by maintaining the sorbent feed rate at a set (Ca/S) ratio. Data prepared for EPRI that was used in developing the cost algorithms in this section are based on dolomite as the sorbent. As the Ca/S ratio increases, the dolomite feed rate, and thus sorbent cost, increases. Solid waste generation also increases with increasing Ca/S ratios.

Many other factors determine the SO_2 control performance of PFBC systems, including combustion temperature, sorbent particle size and reactivity, and feed mechanisms. However, it is beyond the scope of this analysis to account for those factors. An equation was developed from data in Reference 2 to estimate the Ca/S ratio for a specified SO_2 reduction.

$$CAS = EFFSO_2/EFFCA \quad (1)$$

where CAS = Ca/S ratio

$EFFSO_2$ = SO_2 removal efficiency (percent)

EFFCA = calcium utilization efficiency for dolomite (percent).

Reported data show calcium utilization for dolomite to be 60 percent.[2] Therefore, for 90 percent SO_2 removal, the Ca/S ratio is 1.5. The dolomite feed rate can then be calculated by the following equation:

$$DFR = M_{coal} * (\%S/100) * CAS * 100.09/(32.06 * 0.52) \quad (2)$$

where DFR = dolomite feedrate (tons/hr)
%S = percent sulfur in coal
M_{coal} = coal rate (tons/hr), calculated by the following:

$$M_{coal} = (MW * NHR)/(HV * 2) \quad (3)$$

where NHR = PFBC net heat rate (Btu/kW-h)
HV = coal higher heating value (Btu/lb)
MW = net plant megawatt rating.

The amount of solid waste generated at a PFBC site depends on the ash content of the coal, the dolomite feedrate, and the efficiency of the PM control device. The solid waste consists primarily of coal ash, dolomite inerts, and calcium sulfate ($CaSO_4$) which is a reaction product of the calcined dolomite and SO_2. The following equation takes these components into account for PFBC units:

$$SWGR = [0.9 + 0.1 * (EFFPM/100)] * (0.624 * DFR + M_{coal} * (\%S * EFFSO_2 * 2.5/10{,}000 + \%A/100)) \quad (4)$$

where SWGR = solid waste generation rate (tons/hr)
EFFPM = PM control efficiency (percent)
%A = coal ash content (percent).

Equation 4 is based on dolomite composed of 52 percent CaO and a recycle cyclone efficiency of 90 percent.

3.4 CAPITAL COST MODEL

3.4.1 Process Capital Costs

The direct capital costs for a new PFBC plant are based on total plant cost estimates for 250, 500, and 1,300 MW plants.[3,4] Costs are related to boiler size in megawatts using the following power law relationship:

$$COSTS = a\ (MW)^{b} \tag{5}$$

where MW = net boiler rating, megawatts
a = the coefficient
b = the scaling factors.

The process capital costs were back calculated from the total plant cost estimates by eliminating the indirects.[3,4] Based on this approach, the following equation was developed for new PFBC plants to estimate process capital cost:

$$PRCAP = 4{,}109 * (MW)^{-0.251} \tag{6}$$

where PRCAP = process capital costs (1987 $/kW)
MW = net plant capacity (MW).

The Chemical Engineering Plant Cost Index was used to update the costs to 1987 dollars from those presented in References 3 and 4.

Repowered plant process capital costs were estimated using a slightly different approach than followed for the AFBC and IGCC analyses. For PFBC repowering, the existing plant equipment was evaluated under a reuse factor methodology developed by Snow and Maibodi.[6] Under this approach, the present value of each of the existing plant process areas was first estimated. Reuse factors for each process area were then applied to the present value estimate to determine credits against new PFBC process area costs. It was also assumed

that the reusable base plant equipment would be upgraded and refurbished under a life extension effort concurrent with the repowering program. This methodology can be represented in equation form as:

$$(PCC)_{\substack{\text{Repowered Plant}\\ \text{At Final Capacity}}} = (PCC)_{\substack{\text{New Plant At}\\ \text{Final Capacity}}} - \left(\begin{matrix}\text{Existing Reusable}\\ \text{Equipment Credit}\end{matrix}\right) + (\text{Equipment Refurbishment Cost}) \quad (7)$$

The credit for existing reusable equipment was estimated by first identifying major plant sections for a conventional NSPS power plant, determining the percentage of new plant cost represented by each section using EPRI studies[5], and applying an equipment reuse factor for each section. Table 3-1 shows a plant section equipment listing for a conventional NSPS power plant and estimated reuse factors for PFBC repowering. If the existing equipment is well maintained and compatible with the new PFBC hardware, then a high equipment reuse factor is assigned to the process area. If the equipment has been poorly maintained, is not compatible with new hardware, or is replaced, then a low reuse factor is assigned. For example, it was estimated that the steam turbine generator represents 12.7 percent of the process capital cost for a conventional NSPS plant.[5] Since it is assumed that PFBC repowering substantially reuses an existing, well maintained steam turbine generator, the estimated reuse factor for this item is 80 percent. Therefore, 10.2 percent (0.127 x 0.80 x 100) of the equivalent steam turbine generator value at the base plant generating capacity is taken as a credit against the cost for a greenfield PFBC installation. The steam generator (furnace and boiler) is completely replaced with a new bubbling bed combustor and boiler and associated pressure vessel. Therefore, a zero value for the reuse factor is assigned to the steam generator. When all plant sections are evaluated in this way, the available credit for existing reusable equipment is 50.3 percent of the base plant replacement value.

The concurrent refurbishment, demolition, removal, and tie-in costs were estimated at 20 percent of the reusable base plant values. In effect, this reduces the credit for existing reusable equipment by 20 percent, giving a

TABLE 3-1. EXISTING REUSABLE EQUIPMENT AS A PERCENTAGE OF BASE PLANT COSTS FOR PFBC[a]

Plant Section	Percentage[b] of NSPS Plant Costs	Equipment[c] Reuse Factor	Percentage of Base Plant Cost
Civil Engineering	12.9	0.7	9.0
Architectural & Structural	6.0	0.6	3.6
Steam Generator	24.5	0[d]	0
Steam Turbine Generator	12.7	0.8	10.2
Steam/Water Circuit	6.3	0.8	5.0
Solids Receiving/Handling Equipment	1.7	0.8	1.4
Solid Waste Handling	1.2	0.8	1.0
ESP Particulate Removal	6.6	0.7[e]	4.6
Piping, Ductwork, Insulation, Lagging, Auxiliary Equipment	12.6	0.6	7.6
Stack (including lining, lights, etc.)	1.2	0.6	0.7
Electrical Equipment	8.3	0.6	5.0
Control & Instrumentation	2.6	0.6	1.6
Heating, Ventilating, and Air Conditioning	0.6	1.0	0.6
TOTAL			50.3%

[a] Total plant generating capacity is assumed to remain constant after repowering.

[b] Source: EPRI.[5]

[c] Source: Snow and Maibodi[6]; Waddell et al.[7]

[d] The existing steam generator is replaced with the new PFBC boiler and associated pressure vessel.

[e] Assumes that existing particulate control equipment with minor modification will be suitable for final flue gas cleanup following primary and secondary cyclone separators. If there is no existing ESP or fabric filter, use 0.

final equipment reuse credit of 40.2 percent for PFBC repowering. Therefore, repowered PFBC process capital costs are:

$$\text{PRCAP (Repowered PFBC)} = \text{PRCAP (New PFBC)} - 0.402 * \text{PRCAP (New PFBC)} \quad (8)$$

or

$$\text{PRCAP (Repowered PFBC)} = 0.598 * \text{PRCAP (New PFBC)} \quad (9)$$

3.4.2 Indirect Capital Costs

Indirect capital costs are calculated using the EPRI methodology. Table 3-2 presents the procedures to estimate the indirect capital costs. Because information on operating PFBC systems is limited to several pilot scale programs, relatively conservative indirect cost factors are assumed.

3.5 OPERATING AND MAINTENANCE COST MODEL

The direct operating costs associated with PFBC systems include costs for fuel, sorbent, labor, maintenance, water, electricity (parasitic load), and solid waste disposal. The fuel costs for PFBC systems can be estimated using the equations:

$$FC = 8.76 * CF * MW * NHR * CC \quad (10)$$

where FC = annual fuel costs ($/yr)
CF = capacity factor (fraction)
CC = coal cost rate ($/$10^6$ Btu).

Sorbent costs can be determined by:

$$ALC = DFR * CF * 8760 * DSC \quad (11)$$

where ALC = annual limestone cost ($/yr)
DSC = dolomite cost rate ($/ton).

TABLE 3-2. INDIRECT CAPITAL COSTS FOR NEW PFBC PLANTS: EPRI METHODOLOGY[2,4]

(1) Process Capital Costs (PRCAP)	
(2) Total Plant Costs	
General Facilities (GF)	- 10% of PRCAP
+ Engineering and Home Office (EHO)	- 10% of PRCAP
+ Process Contingency	
Case 1 (PcsC)	- 10% of PRCAP
Case 2 (PcsC)	- 0% of PRCAP
+ Project Contingency	
Case 1 (ProjC)	- 30% of (PRCAP + GF + EHO + PcsC)
Case 2 (ProjC)	- 15% of (PRCAP + GF + EHO + PcsC)
+ PRCAP	
Total Plant Costs (TPC)	- PRCAP + GF + EHO + PcsC + ProjC
(3) Royalties	- 0.5% of PRCAP
(4) Preproduction Costs	
1 month fixed operating costs	
+ 1 month variable operating costs (full capacity)	
+ 25% of full capacity fuel cost for 1 month	
+ 2% of Total Plant Investment	
Preproduction Costs (PRD)	
(5) Inventory Capital	- 60 days of fuel and consumable inventory at full load + 0.5% of TPC
(6) Interest during construction	- 6.2% of TPC[a]
(7) Land	- $6,500/acre

[a]Based on 3-year construction period at a 6.1 percent interest rate and 0 percent inflation and assuming that uniform payments occur at the end of the year.

From EPRI operation labor requirements for a 1,000 MW PFBC plant[2] and assuming the same scaling factor as AFBC plants presented in Section 2.5, the following model was developed for this study to estimate operating labor as a function of plant size:

$$OL = 26{,}767 * (MW)^{0.307} \quad (12)$$

where OL = annual operating labor (man-hrs/yr).

Supervisory labor requirements are estimated to be 20 percent of the operating labor costs. Direct labor costs are estimated by applying the appropriate labor cost rates and labor factors. Maintenance costs are estimated as two percent of the total plant costs based on EPRI methodology.

Water requirements for PFBC units are similar to those for equal sized boilers of different design. Based on EPRI estimated water requirements for the 1,000 MW plant[2], the following equation was developed, assuming a linear relationship with plant size:

$$AWC = 6{,}658 * CF * MW * WCR \quad (13)$$

where AWC = annual water costs ($/yr)
WCR = water cost rate ($/1,000 gal).

Solid waste generation cost for new PFBC are estimated by Equation 15:

$$SW = 8{,}760 * CF * SWGR * SWC \quad (14)$$

where SW = annual solid waste disposal cost ($/yr)
SWGR = solid waste generation rate (tons/yr), from Equation 4
SWC = solid waste disposal cost rate ($/ton).

The indirect operating cost include overhead and annualized capital costs. The EPRI procedure should be used to estimate the indirect operating costs.[4]

3.6 KEY TECHNICAL ISSUES

There is currently one utility boiler size PFBC unit (75 MW) under construction in the United States (Tidd) with another one planned for 1995 (Phillip Sporn). These units are demonstration programs designed to refine the technology for widespread use. The following discussion highlights the technology operating performance and efficiency expectations, and the key environmental and engineering issues. The technical and economic issue discussed are based on pilot plant experience. Additional issues will be identified during the demonstration programs. At this time it is not possible to estimate when these issues will be satisfactorily resolved and the technology will be technically ready for widespread application.

3.6.1 Operating Performance and Efficiency

Nominal net heat rates projected for PFBC range from 8,751 to 10,038 Btu/kW-h. Advanced second generation PFBC designs on the drawing boards have overall efficiency objectives of up to 45 percent (equivalent to a net heat rate of 7,584 Btu/kW-h).[8] These designs incorporate a topping cycle in which medium Btu gas from a gasifier module is fired upstream of the gas turbine to bring the turbine inlet temperature up to 2,200°F. Other advanced designs are based on circulating pressurized fluidized bed boilers. Although several laboratory and pilot scale test programs have demonstrated components and systems, a fully integrated, combined cycle facility has not yet operated. With a variable rating between 40 and 80 MW_{th}, the Grimethorpe test facility in England incorporates one of the largest operational PFBC boilers in existence.[9] The Grimethorpe PFBC operated under partial funding from U.S. DOE through the International Energy Agency between 1980 ad 1984. This facility is not fully combined cycle in that it lacks the gas turbine expander. To develop test data, gas turbine performance was simulated using a stationary cascade blade set. The first commercial combined cycle PFBC to achieve operational status will be a 75 MWe unit at Escatron, Spain. The first U.S. combined cycle installation will be the 75 MWe Tidd repowering demonstration

scheduled to begin operation in the fall of 1990.[10] This project is located at American Electric Power's Tidd Station in Brilliant, Ohio and is being partially funded under Round 1 of the U.S. DOE Clean Coal Technology demonstration program. At a steam superheat temperature of 925°F at 1,300 psi and no reheat, the net plant heat rate for Tidd is projected to be 10,038 Btu/kW-h. The Phillip Sporn PFBC repowering is also partially funded under the U.S. DOE Clean Coal Technology demonstration program (Round 2). This project is located at American Electric Power's Phillip Sporn Station near New Haven, West Virginia and is intended to provide commercial operating experience at larger scale. At a rated net capacity of 330 MWe, the Sporn PFBC will incorporate a reheat cycle to give a projected net heat rate of 8,982 Btu/kW-h.[11]

Load following and part-load efficiency (heat rate) is also an operational limitation for bubbling bed PFBC units. Load turndown is primarily accomplished in two ways: by reducing fuel/air ratio and thus decreasing bed temperature, or by reducing bed height and uncovering in-bed tubes. Some designs may also achieve load turndown by slumping portions of a multiple compartment bed. These approaches may have negative implications on emissions performance and boiler efficiency. Bed temperature reductions below 1,500°F causes combustion efficiency to drop with corresponding increase in CO emissions and ash fixed carbon.

Lower bed heights reduce gaseous residence time for SO_2 emissions. Reduction of bed height may uncover heat transfer tubes which may result in reduced boiler efficiency, since the tube bank may contain evaporator superheat, and reheat sections. In addition, since heat transfer performance depends on a complex interaction of temperature, bed density, bubble properties, particle size, and local particle velocities, optimal design of the tube bank configuration and control schemes for part-load operation is difficult to accomplish.

Very little has been reported on turndown performance of PFBC units. The Grimethorpe test facility has demonstrated turndown to approximately 60 percent maximum continuous rating (MCR) through a combination of bed temperature and bed level reduction.[9] Further turndown to 45 percent MCR was

achieved by combining bed temperature reduction with turbine aerodynamic control. For a single bed bubbling boiler, turndown below 40 percent MCR will generally be accompanied by significant heat rate degradation.

3.6.2 Environmental Impacts

Low emissions of SO_2 and NO_x are generally achieved in PFBC units. Removal efficiencies of 90 to 98 percent for SO_2 have been demonstrated with NO_x emissions at about 0.5 pounds per million Btu. However, as in AFBC, SO_2 control during part-load operation may be somewhat reduced. Also, due to long residence times at high temperature and pressure in the PFBC ductwork, SO_3 emissions of 14 to 21 ppm have been measured.[9]

Emissions of N_2O are another concern facing both bubbling and circulating bed PFBCs. Available data are limited, however N_2O emissions at the same order of magnitude as NO (between 50 and 200 ppm) have been reported.[12] These data were obtained with measurement procedures designed to avoid the N_2O measurement artifacts recently identified.

Solid waste production and disposal is another major environmental and economic concerns facing PFBC technology. As higher sulfur fuels are utilized and Ca/S ratios are increased to meet emission standards, the mass of waste material produced increases significantly.[13] In addition to the volume, the spent sorbent is mixed with both the bottom ash and fly ash when it is generally more desirable to produce and collect these streams separately. This mixed waste stream contains reactive lime, reduced sulfur compounds, and carbon, as well as trace metals from the coal ash. Wetting of the material to control dusting creates corrosive calcium hydroxide and substantially increases the mass. Since the ash has been formed at low combustion temperatures, it is not sintered as in the case of PC boiler ash and is therefore more leachable. Finally, although European researchers are investigating the use of the waste as a building material aggregate, the carbon content, increased lime content for high sulfur U.S. coals, leachability issues, and the generally poor pozzolanic qualities may limit this application. Many industrial FBCs are finding waste disposal costs to be significantly higher than projections.

3.6.3 Engineering Issues

Several key issues and problem areas have been identified and must be resolved before industry will have the confidence to commercialize PFBC technology. Some of the issues may represent barriers to widespread application of the technology. Other problem areas are less critical and are similar in scope to engineering problems encountered in the scaleup of any new technology. These engineering problems may serve to delay full commercialization of PFBC in the utility industry or may lead to increased capital and operating costs.

A major problem which remains is the loss of metal from in-bed heat transfer surfaces. Erosion and corrosion phenomena leading to metal loss on submerged in-bed heat transfer surfaces are together termed "metal wastage." Wastage rates of up to 1.6 mm per 1,000 hour of operation have been measured at large scale PFBC demonstration projects.[11] Average wastage rates have been on the order of 0.3-0.5 mm per 1,000 hour. Given a typical heat transfer tube thickness of 5 mm, currently achievable tube bank lifetimes are unacceptable.

Tube bank wastage has been controlled to date through the addition of sacrificial fins and studs to intercept high velocity bubbles and particles, reduction of fluidizing velocity, and modified tube spacing. A fundamental understanding of wastage mechanisms does not exist and, therefore, designs cannot be optimized and tube bank lifetimes cannot be predicted accurately. A potential outcome will be the requirement for more exotic tube materials and reduced fluidization velocities (increased boiler volumes), both affecting capital costs, and more careful selection of fuels and sorbents which impact operating costs.

Erosion and corrosion of turbine blades in the gas turbine has also been a problem in PFBC test programs. To minimize erosion, loadings of less than 0.001 gr/scf are generally required at the gas turbine inlet for particles with diameters greater than 5 microns. This requirement may be difficult to meet with mechanical collectors currently in use at most demonstration sites. Hot gas cleanup systems under development or in proof-of-concept testing include hot ESPs, ceramic candle filters, and

granular bed filters. Testing of a high temperature high pressure (HTHP) ESP at the New York University pilot scale facility was unsuccessful with repeated failure of high voltage insulators. Preliminary proof-of-concept testing of a ceramic candle filter has shown good performance for flowrate of about 8,000 scfm or about 3 MW equivalent.[14] However, scaleup and extended lifetimes are yet to be addressed.

In addition to placing more demands on hot gas cleanup systems, turbine blade erosion and corrosion have been addressed using reduced velocities and pressure ratios through the turbine and advanced metal or ceramic blade claddings. These uncertainties are expected to impact capital cost estimates as well as turbine efficiency and corresponding system heat rate.

Hot metal corrosion has been an even greater problem for combined cycle gas turbines on PFBC units. Alkali sulfidation and chlorine and fluorine attack have been experienced. This problem will increase as turbine inlet temperatures are increased in the more advanced conceptual designs incorporating topping combustors prior to the turbine.

Fuel and sorbent feeding as well as ash removal using rotary valves, lockhoppers, or screw pumps have reliability problems associated with plugging and erosion. Problems have been particularly difficult where feed and removal points are at high temperature and pressure. Again, costs are impacted by the requirement for more expensive materials, greater redundancy, and increased maintenance.

REFERENCES

1. Coal and Syn Fuels Technology. DOE Announces Clean Coal Winners. July 28, 1986. p. 1.

2. Mukherjee, D.K., et al. PFBC Turbocharged Boiler Design and Economic Study. Prepared for the Electric Power Research Institute. Palo Alto, CA. EPRI Report CS-5487, Volume I. November 1987. pp. 5-1 to 5-9.

3. Jensen, A. D., Progress Report on the Tidd PFBC Repowering Project and Comparative Economics of the P-800 PFBC System. In: Comparative Economics of Clean Coal Technologies. Pasha Publications. Arlington, VA. November 1987.

4. Electric Power Research Institute. TAG-Technical Assessment Guide, Volume 1: Electricity Supply - 1986. Palo Alto, CA. EPRI Report P-4463-SR. December 1986. pp. 3-1 to 3-14 and B-52 to B-55.

5. Electric Power Research Institute. Coal-Fired Power Plant Capital Cost Estimates. EPRI Report PE-1865. Palo Alto, CA. May 1981.

6. Snow, G. C., and M. Maibodi. Feasibility Analysis of Advanced Technology Repowering of Coal-Fired Utility Plants. Paper presented at the Joint ASME/IEEE Power Generation Conference, Dallas, TX. October 22-26, 1989.

7. Waddell, J. T., et al. Site-Specific Repowering Options for Fossil Energy Electric Utility Plants. Prepared for the U.S. Department of Energy. Washington, DC. Radian Final Report DCN 89-296-002-01-02. Contract DE-AC01-88FE61666. March 1990.

8. Robertson A. Key Components for a High-Efficiecny Second-Generation Pressurized Fluidized Bed Combustion Combined Cycle. Proceedings: 9th International Conference on Fluidized Bed Combustion. Vol. 2, Boston, MA. May, 1987.

9. Grimethorpe Pressurised Fluidised Bed Combustion Project - Overall Project Review. NCB (IEA Grimethorpe) Ltd. Barnsley, U.K. GEF/U/84/21. May, 1985.

10. Mudd, M.J. Status of AEP's TIDD PFBC Demonstration Plant. Proceedings of the 1989 International Conference on Fluidized Bed Combustion. ASME. New York, NY. Vol. I, pp. 203 - 209. April 20 - May 3, 1989.

11. Rederstorff, B. and Pillai, K.K. Phillip Sporn Plant PFBC Repowering Project. Proceedings of the 1989 International Conference on Fluidized Bed Combustion. ASME. New York, NY. Vol. I, pp. 375 - 380. April 30 - May 3, 1989.

12. Amand, L.E. and Andersson, S. Emissions of Nitrous Oxide (N_2O) from Fluidized Bed Boilers. Proceedings of the 1989 International Conference on Fluidized Bed Combustion. ASME. New York, NY. Vol. I. pp. 49 - 56. April 30 - May 3, 1989.

13. Makansi, J. Users Pause, Designers Wrestle With Fluid Bed Boiler Scaleup. Power. pp. 41 - 42. July 1988.

14. Tassicker, O.J., Burnard, G.K., Leitch, A.J., and Reed, G.P. Performance of a Large Filter Module Utilizing Porous Ceramics on a Pressurized Fluidized Bed Combustor. Proceedings of the 1989 International Conference on Fluidized Bed Combustion. ASME. New York, NY. Vol. I, pp. 479 - 486. April 30 - May 3, 1989.

SECTION 4

INTEGRATED GASIFICATION COMBINED CYCLE (IGCC)

4.1 OVERVIEW

IGCC is an alternative to conventional coal-fired electric power generation with post-combustion emission controls. Emissions of sulfur and nitrogen oxides and particulates from IGCC facilities are projected to be significantly lower than from existing technologies. Figure 4-1 is a generalized block flow diagram of an IGCC facility designed to: (1) convert coal (via partial oxidation and gasification) into a fuel gas, (2) cool and purify the fuel gas, (3) use the clean fuel gas to produce electricity in a combined-cycle combustor turbine application, and (4) treat waste streams generated. The configuration of a specific IGCC facility will depend primarily upon the type of gasification technology used. However, most IGCC facilities will have major process modules similar to those shown in Figure 4-1.

A number of coal gasification processes have been examined for use in combined-cycle applications. The most successful demonstration of IGCC technology in the United States is the Cool Water Coal Gasification Demonstration Project. The Cool Water facility, located near Barstow, California, has operated under the combined sponsorship of EPRI The Synethetic Fuels Corporation, Southern California Edison, and Texaco. The facility is based on Texaco gasification technology and is designed to gasify 1,000 tons of coal per day and produce approximately 100 MW of net electrical power. The IGCC system underwent successful start-up in 1984 to become the first domestic operation combining coal gasification and combustor turbine with heat recovery equipment. The Cool Water facility has exhibited SO_2 removal efficiencies of 96 to 98 percent on both low and high sulfur coals, while reducing NO_x emissions to less than 0.1 lb/10^6 Btu heat input.[1,2]

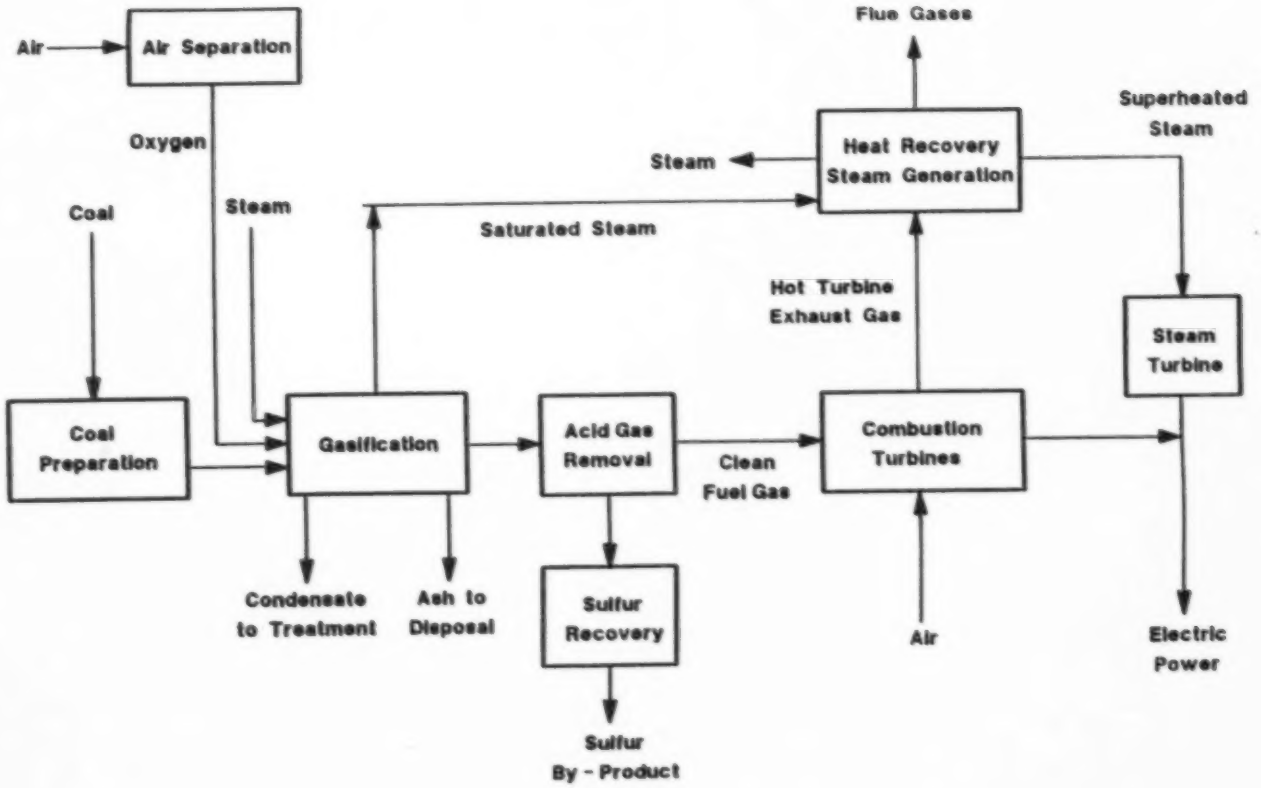


Figure 4-1. Generalized block flow diagram of combined cycle coal gasification power generation.

In this analysis, the cost algorithm for IGCC systems is based on Texaco gasification technology. This design was selected because it has achieved the highest level of demonstration to date. In this design, pulverized coal in a concentrated water slurry is pumped into an entrained flow gasifier. In this partial oxidation process, oxygen is used to produce an intermediate Btu gas (near 280 Btu/scf). An air separation system produces the oxygen required by separating air using heat exchange, Joule-Thompson expansion, and cryogenic distillation. The raw exhaust gas passes through a gas cooling section where particulates are removed by a scrubber. The particulate-free gas is sent to a acid gas removal process where Selexol is used to scrub the sulfur from the gas. The sulfur compounds are reduced to elemental sulfur in a Claus plant. The elemental sulfur is sold as a by-product. The clean gas leaving the acid gas removal process is reheated and is then fired in a combustion turbine with a turbine inlet temperature of about 2200°F. Water injection is used in the gas turbine to reduce NO_x emissions. For the cost models presented in Sections 4.3 and 4.4, the gas turbine is designed for 20°F ambient temperature, without supplementary fuel firing. The hot exhaust gases pass through heat recovery boilers, where steam is generated to drive a steam turbine generator.

A phased approach was assumed in developing the cost algorithms for this design. In Phase I, a combined-cycle, advanced combustion turbine power plant is constructed. This includes an advanced combustion turbine, heat recovery steam generator (HRSG), steam turbine/generator set, cooling system, switchyard, auxiliary power systems, and related systems. The gasification plant, including the gasifier, air separator, and areas for fuel gas cleanup and coal preparation and handling, etc., is built during Phase II. Thus, all costs are separated into the phase where they are incurred.

4.2 IGCC REFERENCE PLANT

Costs have been developed for both new and repowered IGCC systems. A "greenfield" system is defined simply as a new plant which provides new capacity. EPRI cost data were used to develop the model for new and repowered IGCC plants.[3,4] Costs for raw gas cleanup are based on cold gas cleanup

technology. The costs for hot gas cleanup are not included in the cost models because of limited cost data. Costs for a repowered IGCC plant include the cost for any additional equipment or refurbishing of existing equipment necessary to convert a conventional coal-fired power plant to IGCC technology.

4.3 DETAILS OF PHYSICAL PERFORMANCE MODELS

The primary physical characteristic considered in the capital cost model is boiler heat input. Boiler heat input is calculated as a function of net plant output and net heat rate using the following equation:

$$BHI = MW * NHR/1{,}000 \quad (1)$$

where BHI = boiler heat input (10^6 Btu/hr)
MW = net plant output (megawatts)
NHR = net heat rate (Btu/kW-h).

For tail gas treatment and sulfur recovery process areas, the primary characteristic considered in the capital cost model is tons of sulfur removed per day, which is estimated by the following equation:

$$TSR = \frac{12 * MW * NHR}{HV} * \frac{S}{100} * \frac{SO_2EFF}{100} \quad (2)$$

where TSR = tons of sulfur removed (tons/day)
HV = higher heating value of the coal (Btu/lb)
S = sulfur content of the coal (percent)
SO_2EFF = SO_2 removal efficiency (percent).

The costs for both process areas are directly impacted by the amount of sulfur recovered. For the operating and maintenance cost model, coal characteristics (heating value, ash and sulfur content) are taken into account.

The following equation is used to estimate net power output as a function of ambient temperature for the entire IGCC plant:

$$MW = [1.119 - 0.003 * ATEMP] * BMW \quad (3)$$

where MW = net plant output (megawatts)
ATEMP = ambient temperature (^{o}F)
BMW = nominal plant capacity (megawatts).

As discussed in Section 4.1, the capital cost models are based on a gas turbine designed for 20^{o}F ambient temperature without supplementary fuel firing. At higher ambient temperatures, the net plant output will decline because the gas turbine capacity becomes limited. Information from EPRI was used to develop Equation 3 for estimating the net plant output as a function of ambient temperature and nominal (gross) plant capacity.[4] The net power estimated from Equation 3 includes in-plant consumption (i.e., parasitic load). It is assumed that the costs do not vary significantly at different ambient temperatures. From the net power output, capital costs can be estimated on a dollar per kilowatt net basis, and annual costs can be estimated on a mills per kilowatt-hour net basis.

4.4 CAPITAL COST MODEL

4.4.1 Process and Indirect Capital Costs

Table 4-1 presents the process and indirect cost equations developed for new IGCC plants (based on the Texaco design) by process area and by construction phase. As shown by this table, costs are related to boiler heat

TABLE 4-1. PROCESS AND INDIRECT CAPITAL COST EQUATIONS FOR TEXACO'S IGCC PROCESS

Process Area	Phase[a]	Subsystems Included	Process Area Capital Cost Equation 10^6 (1987 $)[b]
Steam, Condensate, and Boiler Feed Water	I	Steam, condensate, and boiler feed water flow scheme	$C = 0.0634 * BHI^{0.6517}$
Gas Turbine and Heat Recovery Steam Generator	I	Gas turbine-generators; heat recovery steam generator	$C = 0.814 * 0.1865 * BHI^{0.7847}$
Steam Turbine	I	Steam turbine-generator	$C = 0.186 * 0.1865 * BHI^{0.7847}$
General Facilities	I	Flare system; tankage; interconnecting piping; cooling water system; raw water supply and wastewater treatment system; fire protection system; electrical system; compressed air system; purge nitrogen system; site improvements; buildings; and switchyard	$C = 2.891 * BHI^{0.4675}$
Coal Handling and Slurry Preparation	II	Coal receiving and storage; coal reclaiming and coal grinding and slurry preparation	$C = 0.0895 * BHI^{0.7135}$
Oxidant Feed	II	Air compression; air cleanup and cooling; high pressure column; low pressure column; and oxygen compression	$C = 0.0466 * BHI^{0.8726}$
Gasification and Ash Handling	II	Coal slurry charging; gasification; steam generation and gas scrubbing; and ash/water separation	$C = 0.2228 * BHI^{0.7643}$

(continued)

TABLE 4-1. (Continued)

Process Area	Phase[a]	Subsystems Included	Process Area Capital Cost Equation 10^6 (1987 $)[b]
Gas Cooling	II	Gas cooling and saturation, and water clarification	$C = 0.0192 * BHI^{0.8477}$
Acid Gas Removal	II	Feed gas chiller; absorber; rich solvent conditioning; and solvent stripping	$C = 0.0415 * BHI^{0.7257}$
Sulfur Recovery	II	H_2S combustion; Claus reaction; sulfur collection; and tail gas incineration	$C = 0.4839 * TSR^{0.6408}$
Tail Gas Treatment	II	Hydrogenation; water quench; and amine absorption	$C = 0.4055 * TSR^{0.5823}$

[a] Phase I is combined cycle and Phase II is gasification.

[b] C = process area capital costs.
BHI = boiler heat input, 10^6 Btu/hour.
TSR = sulfur recovery, tons/day.

input for all process areas except sulfur recovery and tail gas treatment using the following power law relationship:

$$COSTS = a\,(BHI)^{b} \qquad (4)$$

where a = the coefficient
b = the scaling factors.

For the sulfur recovery and tail gas treatment areas, the following equation is used to relate costs to tons of sulfur remove per day:

$$COSTS = a\,(TSR)^{b} \qquad (5)$$

The scaling factors for each process area were determined by relating EPRI cost data for plant sizes of 50, 100, 250 and 1,000 MW to boiler heat input.[3] Using the scaling factors estimated from these data, the coefficient in Equation 4 was calculated based on more recent EPRI data for a 600 MW plant.[4] It should be emphasized that the cost equations derived for each process area include direct costs, indirect costs, engineering, and home office fees expenses. Costs for the combined cycle section (gas turbine with heat recovery steam generator and steam turbine) reported by EPRI were disaggregated into costs for gas turbine with heat recovery steam generator and costs for steam turbine using other EPRI steam turbine cost data.[5] Based on EPRI data, the cost for a steam turbine is 18.6 percent of the costs for the combined cycle section for a 600 MW plant.[5] This percentage is assumed to remain the same for all plant sizes. In addition, a separate cost equation is developed for general facilities.

The process capital cost model for repowered IGCC plants was developed using the following approach:

- For each process area, a reuse factor was assigned based on the percentage of the process area that would not be needed due to reuse of the existing plant. The assignment of the reuse factor was based on existing equipment compatibility with the new hardware as well as general condition of the existing equipment. Reuse factors are very site-specific reflecting differences in plant configuration, age, and maintenance histories. Given sufficient information on a specific plant, a capital cost estimate for repowering can be constructed for that site. Generic reuse factors are estimates of

average equipment compatibility and condition for the population of existing plants suitable for repowering. Using generic reuse factors, a population-wide average capital cost estimate can be prepared for a hypothetical plant. Table 4-2 presents the generic reuse factors for the IGCC process areas. A reuse factor of 1.0 indicates that reuse of the existing equipment will provide full credit or completely offset the need for new equipment within the process area. A reuse factor of 0.0 indicates that there is no equipment suitable for reuse within the process area.

- The reuse factor for each process area was multiplied by the process area cost to obtain a reuse credit.

- It was assumed that reusable equipment will be upgraded and refurbished concurrent with the repowering program. Refurbishment costs were estimated at 20 percent of the replacement cost, thus effectively reducing the weighted reuse credit by 20 percent. Refurbishment costs also include the costs of demolition and removal of some existing equipment.

- The sum of the reuse credits for all process areas is the percent reduction in new IGCC plant capital cost that results from reusing existing plant equipment.

In equation form, the repowered process area capital cost (PRCAP) is:

$$\text{PRCAP (Repowered IGCC)} = \text{PRCAP (New IGCC)} - (1-0.2) * [\text{C (New IGCC)} * \text{Equipment Reuse Factor}] \quad (6)$$

where PRCAP (Repowered IGCC) = total process area capital cost of a repowered IGCC plant (10^6 1987)

PRCAP (New IGCC) = total process area capital cost of a greenfield IGCC plant (10^6 1987)

C (New IGCC) = individual process area capital cost of a new IGCC plant (10^6 1987)

4.4.2 Other Indirect Capital Cost

Other indirect capital costs (such as process and project contingency, royalties, preproduction, inventory capital, etc.) are calculated using EPRI methodology.[4,8] Table 4-3 presents the indirect cost factors for new IGCC plants. The indirect cost factor can be applied to estimate costs for the entire plant for each construction phase.

Because cost data were not available for new IGCC plants firing other coal types, capital costs estimated from the model presented in Section 4.4

TABLE 4-2. EQUIPMENT REUSE FACTOR AS A PERCENTAGE OF NEW IGCC PLANT COST[a]

IGCC Process Area	Equipment Reuse Factor[b,c]
Coal Handling and Slurry Preparation	0.6[d]
Oxidant Feed	0
Gasification and Ash Handling	0
Gas Cooling	0
Acid Gas Removal	0
Sulfur Recovery	0
Tail Gas Treatment	0
Steam, Condensate, and Boiler Feed Water	0.2[e]
Gas Turbine and Heat Recovery	0
Steam Turbine	0.8[f]
General Facilities	0.5[g]

[a]Total plant generating capacity is assumed to remain constant after repowering.

[b]Source: Snow and Maibodi[6]; Waddell et al.[7]

[c]An equipment reuse factor of 0 indicates that the corresponding process area equipment is not present at the existing plant site.

[d]Assumes the existing plant is coal-fired and has existing coal receiving, storage, and crushing facilities that are well maintained and reusable.

[e]Assumes significant new equipment requirements due to increased feed water flow and/or gas turbine wet injection.

[f]Assumes the existing steam turbine and generator are well maintained and are substantially reused.

[g]Assumes additional general facilities are required to support IGCC hardware additions.

TABLE 4-3. INDIRECT CAPITAL COSTS FOR IGCC SYSTEMS: EPRI METHODOLOGY[a]

(1) Process Capital Costs (PRCAP)	
(2) Total Plant Costs	
General Facilities (GF)	= 10% of PRCAP
+ Engineering and Home Office (EHO)	= 10% of PRCAP
+ Process Contingency	
Case 1 (PcsC)	= 10% of PRCAP
Case 2 (PcsC)	= 0% of PRCAP
+ Project Contingency	
Case 1 (ProjC)	= 30% of (PRCAP + GF + EHO + PcsC)
Case 2 (ProjC)	= 15% of (PRCAP + GF + EHO + PcsC)
+ PRCAP	
Total Plant Costs (TPC)	= PRCAP + GF + EHO + PcsC + ProjC
(3) Royalties	= 0.5% of PRCAP
(4) Preproduction Costs	
1 month fixed operating costs	
+ 1 month variable operating costs (full capacity)	
+ 25% of full capacity fuel cost for 1 month	
+ 2% of Total Plant Investment	
Preproduction Costs (PRD)	
(5) Inventory Capital = 60 days of fuel and consumable inventory at full load + 0.5% of TPC	
(6) Interest during construction = 6.2% of TPC[a]	
(7) Land = $6,500/acre	

[a] Based on 3-year construction period at a 6.1 percent interest rate and 0 percent inflation and assuming that uniform payments occur at the end of the year.

could not be compared to costs reported in other sources. Therefore, the sensitivity of the cost model to varying coal types can not be evaluated.

4.5 OPERATING AND MAINTENANCE COST MODEL

Cost models for estimating operating and maintenance (O&M) costs for new IGCC plants are based on EPRI data.[4,8,9] EPRI cost data for the 600 and 400 MW plants were divided into the two IGCC construction phases (combined cycle and gasification) prior to developing the cost model.[4,9] Table 4-4 presents the algorithm for estimating O&M costs for the combined cycle phase, and Tables 4-5 and 4-6 present the algorithm for estimating O&M costs for the gasification phase.

The direct operating costs associated with the entire plant include cost for labor, maintenance, fuel, raw water, catalyst and chemicals, ash/slag disposal, sulfur recovery credit and electricity. Operating labor requirements, in terms of the number of operators per shift, were related to net power output for the 600 and 400 MW plants using EPRI data.[4,9] For the 600 MW plant, operating labor requirements were divided by process area. Therefore, operating labor required for each construction phase could be determined. Because only the total plant labor requirement was presented for the 400 MW plant[9], it was assumed that the percentages of total plant labor required for each construction phase and process area would be the same for both the 400 MW and 600 MW plants. The operating labor cost equations shown in Tables 4-4 and 4-5 include a 4.75 shift breaker factor.[4] This shift breaker factor assumes a 24 hour, 7 days a week coverage with an allowance for sick leave, vacation, etc. The operating labor requirement is assumed to be mandatory regardless of capacity factor.

Maintenance costs were estimated based on a percentage of the total plant cost for each process area.[4] As with the operating costs, the maintenance costs are disaggregated into the two construction phases as shown in Tables 4-4 and 4-6. Maintenance labor and materials are assumed to

TABLE 4-4. OPERATING AND MAINTENANCE (O&M) COST EQUATIONS FOR TEXACO IGCC PROCESS: COMBINED CYCLE PHASE

Item	Subsystem Included	O&M Cost Equation[b]
Operating Labor	Power generator, raw water/ cooling system, instrumentation, and wastewater treatment	OLC = 8,760 * OLR * (7.5 + 0.0025 * MW)
Maintenance[a]	Steam, condensate, and boiler feed water, gas turbines, steam turbines, heat recovery steam generator, general facilities, and switchyard	MC = 1.5% of the total plant cost for combined cycle area including general facilities and switchyard
Administrative And Support Labor		ASC = 30% of O&M labor
Fuel	Operation of combined cycle phase only	FUEL = 8.76 * CF * NHRCC * MWCC * FC
Raw Water	Cooling tower, boiler, demineralizer, and miscellaneous	WTR = 3,571 * CF * NHR * MW * WR/HV
Catalyst And Chemicals	Turbines, cooling tower, wastewater treatment, boiler feedwater treatment	CHEM = 3,662.1 * NHR * MW/HV

[a]Maintenance labor and materials costs are assumed to be 40 to 60 percent of the maintenance costs, respectively.[4]

[b]OLC = operating labor costs ($/yr)
OLR = operating labor rate ($/hr)
MC = maintenance costs ($/yr)
ASC = administrative and support costs ($/yr)
FUEL = fuel costs ($/yr)
FC = fuel price ($/$10^6$ Btu)
CF = capacity factor (fraction)
WTR = raw water cost ($/yr)
WR = raw water cost rate ($/1,000 gal)
HV = higher heating value of coal to be gasified (Btu/lb)
NHRCC = net heat rate of combined cycle phase only (Btu/kW-h)
MWCC = net power output of combined cycle phase only (megawatts)
CHEM = catalyst and chemicals ($/yr).

TABLE 4-5. OPERATING AND MAINTENANCE (O&M) COST EQUATIONS FOR TEXACO IGCC PROCESS: GASIFICATION PHASE

Item	Subsystem Included	O&M Cost Equation[b]
Operating Labor	Gasifier, air separator, raw gas cooling, acid gas removal, sulfur recovery, tail gas treatment, coal preparation and handling, particulate/ash removal	OLC = 8,760 * OLR * (12.5 + 0.0025 * MW)
Maintenance[a]	Percentages of the total plant cost for all related process areas (See Table 4-6)	
Administrative And Support Labor		ASC = 30% of the O&M labor
Coal Consumption		CFUEL = 8.76 * CF * NHR * MW * CC
Raw Water	Gasifier, sulfur recovery, coal handling, wastewater treatment, and biotreatment	WTR = [-1,577 + (1,380 * NHR * MW/HV)] * CF * WR
Catalyst And Chemicals	Gasifier, wastewater treatment, biotreatment, acid gas removal, sulfur recovery, and tail gas treatment	CHEM = (3,946 * NHR * MW/HV) + (78.05 * %S * NHR * MW/HV) * SO_2EFF * SEFF/10,000
Ash/Slag Disposal	Gasifier, particulate/ash removal and wastewater treatment	AD = 43.8 * (CF * NHR * WDR * MW/HV) * (%C/100 + %A)

(continued)

TABLE 4-5. (Continued)

Item	Subsystem Included	O&M Cost Equation[b]
Sulfur Recovery Credit	Sulfur recovery (Claus Process)	SC = (43.8 * %S * NHR * MW/HV) * CF * SO_2EFF * SEFF * SCR/10,000

[a] Maintenance labor and materials costs are assumed to be 40 and 60 percent of the maintenance costs, respectively.[4]

[b] CFUEL = coal costs ($/yr)
CC = coal price ($/$10^6$ Btu)
%S = sulfur content of the coal, as received (wt %)
SO_2EFF = overall SO_2 removal efficiency (%)
SEFF = overall sulfur recovery (%)
AD = ash disposal costs ($/yr)
%C = carbon content of the coal, as received (wt %, if unknown assume 61.2%)
%A = ash content of the coal, as received (wt %)
WDR = waste disposal costs ($/ton)
SC = sulfur recovery credit ($/yr)
SCR = sulfur selling price ($/ton).

TABLE 4-6. MAINTENANCE COSTS AS A FRACTION OF TOTAL PLANT COST FOR GASIFICATION PHASE

Process Area	Percent of Total Plant Cost
Coal Handling and Slurry Preparation	3.0
Oxidant Feed	2.0
Gasification and Ash Handling	4.5
Gas Cooling	3.0
Acid Gas Removal	2.0
Sulfur Recovery	2.0
Tail Gas Treatment	2.0

be 40 and 60 percent of the maintenance costs, respectively.[4] Administrative and support labor costs are estimated at 30 percent of the operating labor costs.[4]

Fuel costs for a new IGCC plant correspond to the amount of coal consumed by the process. If only the combined cycle is operating, fuel costs would be based on firing either natural gas or distillate oil in the gas turbine. Regardless of which fuel is fired, fuel costs are estimated using the equations shown in Tables 4-4 and 4-5. The raw water cost equations in Tables 4-4 and 4-5 were derived from EPRI water requirements for the 600 MW plant[4] and for the 400 MW plant.[4] From detailed material balances presented in both references, water requirements in gallons per hour for each construction phase were calculated. The water requirements were linearly related to coal consumption rate. Using the equations in Tables 4-4 and 4-5, raw water costs are obtained by multiplying the water requirement by the appropriate water cost rate and the hours of operation.

A similar approach was used to develop the equations shown in Tables 4-4 and 4-5 for catalyst and chemicals, ash disposal, and sulfur recovery credit using detailed Radian Corporation cost information for the 600 MW plant.[4] The catalyst and chemical cost equation for the combined cycle phase in Table 4-4 was derived by relating the costs to the coal consumption rate for the 600 MW plant. The catalyst and chemical costs for the gasification construction phase were further subdivided into costs depending only on size (related to coal consumption rate) and costs depending on sulfur removal rate. The costs which depend upon sulfur removal rate are those chemicals required for acid gas removal, sulfur recovery, and tail gas treatment processes. Therefore, the equations shown in Table 4-5 are based on size (i.e., coal consumed) and the amount of sulfur captured. The EPRI IGCC plants are based on 95 percent SO_2 removal and 99.8 sulfur recovery efficiencies.[4] The sulfur recovery credit equation shown in Table 4-5 is based on the amount of sulfur recovered (for later sale) from that in the coal for the 600 MW plant. The equation for ash/slag disposal in Table 4-5 is based on the amount of ash in the coal and the amount of unburned carbon for the 600 MW plant. The combustion efficiency used to estimate unburned carbon is 99 percent.[4] The capital costs for IGCC are annualized using procedures developed by EPRI.[8]

4.6 KEY TECHNICAL ISSUES

Several integrated coal gasification power generation cycles have been demonstrated in the U.S. Many of the major technical issues have been resolved for designs having commercially available cold gas clean-up systems. Wider spread application of this technology will be a function of coal to natural gas fuel price differential, improved system heat rate, and overall capital cost reductions. Current coal-to-natural gas fuel price differential favors the installation of gas combustion turbines versus conventional or advanced coal technologies. Although heat rate improvements can be achieved through more efficient heat recovery designs, much greater improvements can be achieved by development of hot gas cleanup systems and higher temperature

combustor turbines. However, the time frame for successful development of hot gas clean-up systems is unknown. Utilities are currently installing significant new capacity in gas/oil combustion turbines which can be modified to burn coal derived gas. This makes coal gasification a viable retrofit technology for these combustion turbines should the price differential between coal and gas increase significantly. Until the coal to gas fuel price differential significantly increases, IGCC will not become widely used unless hot gas clean-up systems and high temperature gas turbines become technically and economically feasible.

4.6.1 Operating Performance and Efficiency

Integrated gasification combined cycle power generation systems have the potential to generate electricity at efficiencies of at least 40 percent while controlling environmental emissions to extremely low levels. In certain configurations, by-products, including elemental sulfur, methanol, and syn-gas, may be produced and sold for credit, thus reducing the overall levelized cost of electricity.

The most efficient conceptual design considered to be commercially available on a near-term basis appears to be the British Gas Corporation/Lurgi slagging gasifier with hot gas cleanup and a gas turbine inlet temperature of 2,200°F.[10] The projected heat rate for this design is 8,661 Btu/kW-h.[8] Other configurations utilizing current technology gas turbines with inlet temperatures at about 2,000°F give about a 6 percent heat rate degradation to about 9,200 Btu/kW-h.[8,10] For systems incorporating quench gas cooling and current turbine technology, projected heat rates are between 9,960 and 10,000 Btu/kW-h.[5,12]

The Cool Water demonstration project, at a nominal capacity of 100 MW, operates at an average heat rate of 10,600 Btu/kW-h. Although this facility is in commercial operation, its heat rate is relatively high (i.e., overall generating efficiency of 32 percent). However, Cool Water is considered to be a first-of-a-kind facility, having as its primary objectives the testing and

evaluation of individual components. Second-of-a-kind (beyond Cool Water) facilities are expected to be further optimized to give improved heat rates.

Projected net plant heat rate depends significantly on gasifier type, gas cleanup process (hot or cold), and gas turbine efficiency (primarily a function of turbine inlet temperature). Reported heat rate is also influenced by ambient temperature. Development towards simplified, more fully integrated systems is intended to match temperature and pressure characteristics through the system. Major improvements in heat rate can be expected by eliminating quench cooling steps, and increasing turbine inlet temperature. These efforts, however, depend on further breakthroughs in hot gas cleanup, including particulate, sulfur, nitrogen, and alkali removal.

Load following and turndown for IGCC plants is generally accomplished in two ways: turbine following and turbine lead. Under the turbine following approach, turndown is determined by gasifier control. The maximum expected turndown per gasifier module is expected to be between 50 and 60 percent of maximum continuous rating (MCR). For a two module system the maximum turndown would therefore be approximately 25 percent MCR. This would be accompanied by about a 13 percent degradation in heat rate.[5] Under the turbine lead (gasifier following) approach, the maximum turndown per turbine is approximately 75 percent MCR. For a two turbine plant the maximum turndown would be approximately 38 percent MCR with a corresponding heat rate degradation of about 10 percent.[5]

4.6.2 Engineering and Environmental Issues

Major issues facing IGCC technology are primarily associated with hot gas sulfur and ammonia removal. In addition, further development is required to improve hot gas particulate control for protection of external sorbent beds and the gas turbine. Alkali corrosion of the gas turbine blades will also become more serious at increasing inlet temperature.

Issues that must be overcome before IGCC is considered fully viable are:

- In-Situ Sulfur Capture
  - Attempt to achieve up to 90% sulfur capture within the gasifier reacting bed
  - Requires in-bed sorbent
  - Waste volume and reactivity are concerns
- External Desulfurization
  - Fixed, moving, or fluidized bed vessels
  - Selexol requires gas quench to 100°F
  - Zinc ferrite
    + Limited to 900-1,200°F
    + Poisoned by carbon carry-over from gasifier
    + Long term demonstration of sulfidation and regeneration required (fewer than 20 cycles have been demonstrated to date)
  - Lime/limestone operational to 1,800°F
  - Advanced metal oxides (iron oxide, sodium oxide) under development for temperatures to 1,350°F
  - Requires hot particulate control prior to sorbent bed
  - Requires improved sorbent mechanical durability (crush strength)
- Sorbent/Sulfur Disposition
  - Stability, leachability, and reactivity for direct disposal are issues
  - May require "bottom ash oxidizer" to oxidize sulfided sorbent to sulfates
  - Sorbent regeneration may have major impact on system economics
  - Regeneration efficiency decreases with number of cycles
- Particulate Removal
  - Hot cyclones
    + Demonstrated to 1,600°F

    + Cleanup to 0.1 gr/scf
    + Inefficient part-load operation

- Hot ESP
    + High temperature/high pressure operation not demonstrated
    + Insulator integrity a concern

- Barrier filters
    + Under development
    + Temperature potential to 1,800°F
    + Ceramic candle filters
    + Ceramic cross flow filters

- NO_x Control

    - NH_3 removal from gasifier product gas difficult at high temperature

    - Thermal NO_x from combustion turbine
        + Low NO_x combustors
        + Wet injection

- Demonstration of Advanced (High Temperature) Gas Turbines

REFERENCES

1. Clark, E., Cool Water Coal Gasification Programs, an Update. EPRI Journal. 88:45-48, September 1988.

2. Synthetic Fuels Associates, Inc. Coal Gasification Systems: A Guide to Status, Applications, and Economics. EPRI Report AP3109. June 1983.

3. Fluor Engineers, Inc. Economic Assessment of the Impact of Plant Size on Coal Gasification-Combined-Cycle Plants. Prepared for the Electric Power Research Institute. Palo Alto, CA. EPRI No. AP-3084. May 1983.

4. Bechtel Group, Inc. An Evaluation of Integrated-Gasification-Combined Cycle and Pulverized-Coal-Fired Steam Plants. Volume 1: Base Case Studies. Prepared for the Electric Power Research Institute. Palo Alto, CA. EPRI No. AP-5950. September 1988.

5. Bechtel Group, Inc. Capital Cost Estimates and Schedules for Coal-Fired Power Plants. Prepared for the Electric Power Research Institute. Palo Alto, CA. EPRI No. P-4542. August 1986.

6. Snow, G. C., and M. Maibodi. Feasibility Analysis of Advanced Technology Repowering of Coal-Fired Utility Plants. Paper presented at the Joint ASME/IEEE Power Generation Conference, Dallas, TX. October 22-26, 1989.

7. Waddell, J. T., et al. Site-specific Repowering Options for Fossil Energy Electric Utility Plants. Prepared for the U.S. Department of Energy. Washington, DC. Radian Final Report DCN 89-296-002-01-02. Contract DE-AC01-88FE61666. March 1990.

8. Electric Power Research Institute. TAG-Technical Assessment Guide, Volume 1: Electricity Supply - 1986. Palo Alto, CA. EPRI Report P-4463-SR. December 1986.

9. Bechtel Group, Inc. An Evaluation of Integrated-Gasification-Combined-Cycle and Pulverized-Coal-Fired Steam Plants. Volume 2: Sensitivity Studies and Appendices. Prepared for the Electric Power Research Institute. Palo Alto, CA. EPRI No. AP-5950. September 1988.

10. Holt, N.A. and Gluckman, M. Integrated Coal Gasification Combined Cycle Plants--A Flexible Generation Planning Strategy. Proceedings: First International Conference on Improved Coal-Fired Power Plants. EPRI. Palo Alto, CA. CS-5581-SR. pp. 2-277 - 2-314. 1988

11. delaMora, J.A., et al. Evaluation of the British Gas Corporation/Lurgi Slagging Gasifier in Gasification-Combined-Cycle Power Generation. EPRI No. AP-3980. March 1985.

12. Hartman, J.J., et al. Shell-Based Gasification-Combined Cycle Power Plant Evaluations. EPRI No. AP-3129. June 1983.

SECTION 5

PULVERIZED COAL-FIRED PLANT

5.1 OVERVIEW

Pulverized coal combustion is a mature technology that has been developed over the last 100 years for electrical production. Pulverized coal power plants will continue to be used in the future. However, these plants will require add-on SO_2 and NO_x controls to meet stringent environmental regulations.

This section presents detailed cost estimating procedures for pulverized coal-fired power plants. The costs for a baghouse for PM emission control to 0.03 $lb/10^6$ Btu as well as for FGD and NO_x controls are included from other sources. The costs for a pulverized coal power plant estimated from the model developed in this study can be added to the costs for add-on PM, NO_x and SO_2 controls to arrive at total estimated costs for a plant required to meet stringent environmental regulations. The total plant costs for the pulverized coal plant with add-on controls can be compared with those for more advanced steam generating processes (FBC and IGCC) that are also designed to achieve low SO_2 and NO_x emissions while generating power.

5.2 PC REFERENCE PLANT

Itemized costs are estimated for new plants firing a Wyoming subbituminous and a high sulfur Midwestern bituminous coal located at Kenosha, Wisconsin.[1,2] A new plant is defined simply as a grassroots plant that provides new capacity. Costs are estimated based on the following process areas: steam generator and stack, PM controls (baghouse and ID fan), waste handling and disposal, steam/boiler feed water/condensate system, steam turbine generator, coal handling and preparation, and general facilities (buildings, site development, electrical equipment, and distributables).

5.3 DETAILS OF PHYSICAL PERFORMANCE MODELS

Fuel characteristics and boiler heat input are the main factors affecting pulverized coal-fired plant costs. These factors were incorporated into the cost models for estimating costs for plants firing other coal types besides Wyoming subbituminous and Illinois bituminous coals. For most process areas, boiler heat input is the only factor used in the cost model. Boiler heat input is estimated by the following equation:

$$BHI = \frac{NHR * MW}{1000} \quad (1)$$

where BHI = boiler heat input, 10^6 Btu/hr
NHR = net heat rate, Btu/kW-h
MW = net plant output, megawatts.

For PM controls and coal handling, the coal rate is used as the basis for the cost model. Coal rate is calculated as:

$$CR = BHI/HV \quad (2)$$

where CR = coal rate, 10^6 lb/hr
HV = higher heating value of coal (as received), Btu/lb.

For waste disposal capital and annual costs, the ash rate in terms of millions of pounds per hour (10^6 lb/hr) is used as the basis for the cost models.

5.4 CAPITAL COST MODEL

5.4.1 Process Capital Costs

The process capital cost equations for a new pulverized coal-fired plant are shown in Table 5-1. The following approach was used to develop the

TABLE 5-1. PROCESS CAPITAL COST EQUATIONS FOR PULVERIZED COAL-FIRED POWER PLANTS

Process Area	Equipment Included	Process Area Capital Cost Equation, 10^6 (1987$)[a]
Steam Generator	Steam generator and stack	$C = 0.097 * (BHI)^{0.799} + 9.11 * (CR)^{0.637}$
PM Controls	Baghouse and ID fan	$C = 51.95 * (CR)^{0.878} - 9.11 * (CR)^{0.637}$
Waste Handling/ Disposal[b]	Ash handling	$C = 0.041 * (ASH)^{0.772}$
Steam/Boiler Feed Water/Condensate	Condenser, rotating equipment, heater and exchangers, tanks and vessels, water treatment/ chemical feed, piping, and control/instrumentation	$C = 0.079 * (BHI)^{0.787}$
Steam Turbine	Steam turbine generator	$C = 0.033 * (BHI)^{0.854}$
Coal Handling/ Preparation	Coal receiving/handling, storage, and conveying	$C = 17.31 * (CR)^{0.480}$
General Facilities	Concrete, structural architectural and finish, earthquake, piles, site preparation, electrical equipment, bulk materials, switch-yard, and miscellaneous.	$C = 0.154 * (BHI)^{0.813}$

[a] C = Process area capital costs;
ASH = Ash production rate, tons per day calculated using
ASH = 0.12 * Ash % * NHR * MW/HV.
Ash % = percent of ash in the coal.

[b] Excludes waste production equipment used in the FGD unit.

capital cost equations. Process area capital cost data for the 1 x 330, 2 x 330, and 2 x 500 MW power plants from Reference 1 were used to develop cost equations based on the following power law relationship:

$$\text{Cost} = a * (\text{variable})^{b} \qquad (3)$$

where a = constant
b = scaling factor
variable = process or design parameter.

The variable used to estimate the costs for different sizes was chosen to reflect coal type sensitivity. For example, the steam generator cost equation in Table 5-1 was based on boiler heat input. Boiler heat input varies for the same boiler firing different coals at the same coal rate.

The costs estimated from the cost equations (updated to 1987 dollars using the Chemical Engineering Plant Cost Index) were compared to recent EPRI costs for a 1 x 200 and a 2 x 500 MW power plant.[2] If the agreement of the costs was not within ±20 percent, the constant, a, was adjusted based on the EPRI cost data for the 1 x 200 MW plant.[2] This approach assumed that the scaling factor, b, is essentially the same for EPRI-reported costs.[1,2]

5.4.2 Indirect Capital Costs

Table 5-2 presents the indirect costs based on the EPRI methodology.[3] The percentages selected for engineering and home office fees, process contingencies are based on those used by EPRI.[2] However, the other indirect costs such as a 30 percent project contingency are based on other EPRI methodologies.[3]

Using the cost model described in Section 5.4.1, capital costs were compared to EPRI costs for the 1 x 200 MW plant firing different coal types.[2] A 0 percent process contingency and a 15 percent project contingency was used in the cost model for this comparison[2] Table 5-3 summarizes the results of this comparison. The costs predicted by the model were at the most

capital cost equations. Process area capital cost data for the 1 x 330, 2 x 330, and 2 x 500 MW power plants from Reference 1 were used to develop cost equations based on the following power law relationship:

$$\text{Cost} = a * (\text{variable})^{b} \qquad (3)$$

where a = constant
b = scaling factor
variable = process or design parameter.

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Table 5-2. INDIRECT CAPITAL COSTS FOR PULVERIZED COAL-FIRED PLANTS: EPRI METHODOLOGY[2,3]

(1) Process Capital Costs (PRCAP)	
(2) Total Plant Costs	
General Facilities (GF)	= 10% of PRCAP
+ Engineering and Home Office (EHO)	= 10% of PRCAP
+ Process Contingency	
Case 1 (PcsC)	= 10% of PRCAP
Case 2 (PcsC)	= 0% of PRCAP
+ Project Contingency	
Case 1 (ProjC)	= 30% of (PRCAP + GF + EHO + PcsC)
Case 2 (ProjC)	= 15% of (PRCAP + GF + EHO + PcsC)
+ PRCAP	
Total Plant Costs (TPC)	= PRCAP + GF + EHO + PcsC + ProjC
(3) Royalties	= 0
(4) Preproduction Costs	
1 month fixed operating costs	
+ 1 month variable operating costs (full capacity)	
+ 25% of full capacity fuel cost for 1 month	
+ 2% of Total Plant Investments	
Preproduction Costs (PRD)	
(5) Inventory Capital = 60 days of fuel and consumable inventory at full load + 0.5% of TPC	
(6) Interest during construction = 6.2% of TPC[a]	
(7) Land = $6,500/acre	

[a]Based on 4-year construction period at a 6.1 percent interest rate and 0 percent inflation and assuming that uniform payments occur at the end of the year.

TABLE 5-3. COMPARISON OF THE PREDICTED AND REPORTED CAPITAL COST FOR A 200 MW PLANT

Plant Location	Coal Type	Total Capital Requirement, $/kW (1988$) Reported with FGD[a]	Predicted with FGD[b]	Percent Error[c]
Kenosha, WI	Illinois No. 6 Bituminous	1,680	1,690	0.6
Bethlehem, PA	West Virginia Bituminous	1,733	1,678	-3.2
Kenosha, WI	Wyoming Subbituminous	1,652	1,713	3.7
Velva, ND	North Dakota Lignite	1,737	1,754	1.0
Freeport, TX	Texas Lignite	1,692	1,751	3.5

[a]Cost reported in Reference 2 i 1982$ and updated to 1988$ for presentation here.

[b]Costs for FGD from EPRI[2] were added to the cost estimated from the capital cost model presented in this section.

[c]Percent Error = 100 * (Predicted with FGD - Reported)/Reported.

3.7 percent different than those reported to EPRI.[2] Some of these cost differences a attributable to diverse plant locations where labor rates and productivity differed. Costs for FGD systems were estimated from EPRI data.[2] Equations for FGD costs were derived using the same approach as described in Section 5.4.1.

5.5 OPERATING AND MAINTENANCE COST MODEL

The direct operating costs associated with pulverized coal-fired plants include costs for fuel, labor, maintenance, water, and solid waste disposal. The fuel costs can be estimated using the equations:

$$FC = 8.76 * CF * MW * NHR * CC \quad (4)$$

where FC = fuel costs ($/yr)
CF = capacity factor (fraction)
CC = coal cost rate ($/$10^6$ Btu).

Based on EPRI information[2], a model for coal-fired power plants was developed for this study to estimate operating labor for these plants, excluding the labor requirements for FGD estimated for EPA[4], as a function of plant size:

$$OL = 37{,}120 * (MW)^{0.258} \quad (5)$$

where OL = annual operating labor (man-hrs/yr).

The operating labor required for a plant without a baghouse is as follows:

$$OL = 37{,}120 * (MW)^{0.258} - (3.51 * ACFM^{0.4} + 0.00146 * ACFM) \quad (6)$$

where ACFM = flue gas flow rate at air preheater exit (ft^3/min).

The baghouse labor requirements are estimated from an EPA equation and were subtracted from those estimated by Equation 5.[4] Operating labor cost is estimated as the product of the annual labor requirement and the labor cost rate. Supervisory labor requirements were estimated to be 20 percent of the operating labor requirements. Total maintenance costs were estimated as two percent of the total plant costs based on EPRI methodology. Maintenance labor and materials are 40 and 60 percent of the total maintenance costs, respectively.[3]

Water requirements for makeup water in the steam cycle were estimated from EPRI data.[2,5] An EPRI report was used to estimate the amount of water needed for the FGD system compared to the makeup water.[5] This ratio of FGD to makeup water was applied to the total water required for the plant/FGD unit for estimating makeup water required only for the plant. The plant makeup water requirement was related to the EPRI coal rate for the 1 x 200 and 2 x 500 MW plants.[2] The following equation was developed to estimate water costs:

$$AWC = (-3,793 + 3,548,900 * CR) * WCR * CF \quad (7)$$

where AWC = annual water costs ($/yr)
WCR = water cost rate ($/1,000 gal).

Solid waste generation cost were estimated as:

$$SW = 365 * CF * ASH * SWC \quad (8)$$

where SW = annual solid waste disposal cost ($/yr)
ASH = solid waste generation rate (tons/day) from equation in Table 5-1
SWC = solid waste disposal cost rate ($/ton).

For plants without baghouses, the solid waste generation rate (ASH) is the amount of bottom ash collected by the boiler. The bottom ash rate is roughly estimated as 20 percent of the solid waste generation rate (ASH) from

the equation in Table 5-1. It is assumed that of the total ash in the coal, 80 percent of the ash leaves the boiler with the flue gas.

Indirect operating costs include overhead and annualized capital costs. An EPRI procedure was used to estimate these indirect operating costs.[3]

5.6 KEY TECHNICAL ISSUES

The conventional PC boiler is a fully mature technology with decades of design and operational experience in the industry. Most of the key engineering issues have been addressed. Continued improvement efforts are focused on refinements in heat rate, plant availability, and monitoring and control systems. Emission control equipment including FGD systems, boiler and duct sorbent injection, low NO_x firing systems, selective catalytic and non-catalytic reduction systems, ESP and fabric filter systems are applicable to PC boilers in both retrofit and new construction configurations. Net station heat rates for conventional PC-based plants range from 9500 to 12,500 Btu/KW-h with an industry median of 10,600 Btu/kW-h.

Critical and supercritical boilers also utilize PC firing systems and one or more of the emission control systems mentioned above. The performance and issues associated with these emission control technologies will not be further discussed here. A critical boiler incorporates a Rankine steam cycle operating at or above the critical pressure of water, which is 3208 psia. Achievable net station heat rates are less than 8500 Btu/kW-h for a supercritical plant.

5.6.1 Operating Performance and Efficiency

As early as 1962, Philadelphia Electric Company's Eddystone No. 1 supercritical unit (at Eddystone, Pennsylvania) demonstrated an annual net plant heat rate of 8534 Btu/kW-h.[6] This plant had a throttle pressure of 5000 psig, a superheat temperature of 1200°F, and a second reheat temperature of 1050°F. This plant is not equipped with an FGD scrubber system which, due to about a 2 percent increase in parasitic power consumption[7], will increase the

net plant heat rate to about 8705 Btu/kW-h. A proposed 4500 psig, 1100°F, 1050°F, 1050°F cycle incorporating state-of-the-art boiler and steam turbine designs as well as an FGD scrubber could achieve a net plant heat rate of 8335 Btu/kW-h.[6]

5.6.2 Engineering Issues

The following key issues and problem areas are being addressed in an effort to make advanced pulverized coal heat cycles more acceptable to utilities:

- Tube bank problems caused by tube metal exfoliation, high-temperature gas side corrosion, high-temperature creep failure, and long-term overheat;
- Steam turbine blade erosion due to tube metal exfoliation, water chemistry, turbine steam control valve cracking, and high-temperature expansion;
- Steam line design for high-temperature expansion and creep failure;
- Load following and control resulting from variable pressure, full ARC admission, and high-capacity bypass of reheat steam; and
- Critical pressure letdown during turbine trip minimizing noise and pressure and temperature decay following steam turbine trip.

From the technical issues and problems presented above, higher capital costs will be required for boilers with advanced pulverized coal heat cycles compared to conventional boilers. The tube banks would be made of better metal alloys and would have thicker walls to withstand gas-side corrosion, high pressure, and creep failures. In addition, the steam turbine blade would be made of better metal alloys to withstand the increased potential for erosion. Operating labor and maintenance costs would also increase from those

of conventional boilers because of the higher temperature and pressure operation. However, fuel cost would be lower because of the lower heat rates for the advanced pulverized coal heat cycles.

REFERENCES

1. Bechtel Group, Inc. Capital Cost Estimates and Schedules for Coal-Fired Power Plants. Prepared for the Electric Power Research Institute. Palo Alto, CA. EPRI Report P-4542. August 1986. pp. 4-1 to 4-2.

2. Bechtel Group, Inc. Evaluation of Alternative Steam Generator Designs for Atmospheric Fluidized Bed Combustion Plants. Prepared for the Electric Power Research Institute. Palo Alto, CA. EPRI Report CS-5296. July 1987.

3. Electric Power Research Institute. TAG-Technical Assessment Guide. Volume 1: Electricity Supply-1986. Palo Alto, CA. EPRI Report P-4463-SR. December 1986. pp.3-1 to 3-14.

4. Palmisano, P.J., and B.A. Laseke. Users's Manual for the Integrated Air Pollution Control System Design and Cost-Estimating Model: Version II, Volume I. EPA-600/8-86-031a (NTIS PB87-127767). U.S. Environmental Protection Agency. Research Triangle Park, NC. September 1986. p. 134.

5. Bechtel Group, Inc. An Evaluation of Integrated-Gasification-Combined-Cycle and Pulverized-Coal-Fired Steam Plants. Volume 1: Base Case Studies. Prepared for the Electric Power Research Institute. Palo Alto, CA. EPRI Report P-5950. September 1988. Chapters 4 and 5.

6. Bannister, R. L., Silvestri, Jr., G. J., and Hizume, A. Evolution and Performance of High-Temperature and Supercritical Steam Turbines. Proceedings: First International Conference on Improved Coal-Fired Power Plants. pp. 3-209 to 3-229. EPRI CS-5581-SR. Palo Alto, CA. 1988.

7. Shattuck, D. M., et al. Retrofit FGD Cost-Estimating Guidelines. EPRI CS-3696. Palo Alto, CA. October, 1984.